Second Annual Clean Coal Technology Conference

Proceedings

Volume 1 September 8, 1993

Objective

This Conference, co-sponsored by the U.S. Department of Energy (U.S. DOE) and the Southern States Energy Board (SSEB), seeks to examine the status and role of the Clean Coal Technology Demonstration Program (CCTDP) and its projects. The Program will be reviewed within the larger context of environmental needs, sustained economic growth, world markets, user performance requirements and supplier commercialization activities. This will be accomplished through in-depth review and discussion of factors affecting domestic and international markets for clean coal technology, the environmental considerations in commercial deployment, the current status of projects, and the timing and effectiveness of transfer of data from these projects to potential users, suppliers, financing entities, regulators, the interested environmental community and the public.

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Plenary Session 1

Moderator:

Jack S. Siegel,
Acting Assistant Secretary for Fossil Energy,
U.S. Department of Energy



Welcome to 2nd Annual Clean Coal Technology Conference Kenneth J. Nemeth September 8, 1992

On behalf of the Southern States Energy Board and the U.S. Department of Energy, it is my privilege to welcome each of you to this Second Annual International Clean Coal Technology Conference here in Atlanta, Georgia. As you glean information from the conference program over the next few days, I hope you will also take time to enjoy our dynamic Olympic city.

A clear understanding of state, regional, national and international issues is no longer peripheral to electricity generation and transmission...it is fundamental to the success of business and government operations. The objective of this conference is to examine the status of the Clean Coal Technology Demonstration Program and its projects. The program will be reviewed within the larger context of environmental needs, sustained economic growth, world markets, user performance requirements and supplier commercialization activities.

Program review will be accomplished through in-depth discussions of factors affecting domestic and international markets for clean coal technology, the environmental considerations in commercial deployment, the current status of projects, and the effectiveness of data transfer to potential users, suppliers, financing entities, regulators and the interested environmental community.

As environmental priorities and energy demands realign themselves, coal emerges as one of the most important energy resources we have here in the United States. Finding new programs that are both innovative and challenging, such as the Clean Coal Technology Demonstration Program,

will allow us to continue to fully utilized our most abundant natural resource, coal.

Last night, many of you attended a tour of Plant Yates. We in the South are very proud that The Southern Company is participating in the Clean Coal Technology Demonstration Program. In fact, the

I think we have prepared a program which accomplishes these objectives, and it is my hope that you will find these next few days in Atlanta fruitful. If I or the SSEB staff can do anything to enhance your stay in Atlanta, please be sure to let us know.

WELCOMING REMARKS

Lee Conn
Vice President Power Generation
Georgia Power Company

(The comments of Mr. Conn were not available at the time of publication.)

Remarks of Deputy Secretary of Energy William White 2nd Annual Clean Coal Technology Conference Atlanta, GA September 8, 1993

Think with me about this. In a sense, here we are in the very middle of a quiet and unsung economic revolution. We are in the very middle of it. Think about this. We are sitting right now the international economy growing and becoming integrated like it never has in the history of the planet.

This will not be the last year that we have international delegations at this conference. They will grow. And when conferences are held abroad it will be <u>Americans</u> who will attend. That's because, as ideologies are swept aside, the common problems of economic growth and the practical problems of building the infrastructure needed to power that growth are things which we will share. That wasnt's true 50 years ago, it wasn't true 100 years ago, it wasn't true 500 years ago. But it is true today.

So when our grandchildren attend conferences like this we won't be recognizing international delegations because it will be taken for granted that conferences about the leading edge of technologies are at the very heart of economic growth throughout the world.

And we're here, frankly, right at the beginning. Now that's exciting.

Maybe I've overblown the topic, but when you think about it, there's something there, and it's not just the revelation of having an economic integration happening before our very eyes. We're seeing -- in the last 10 years and increasingly I predict in the next 10-20 years -- some fundamental redirection in the attitudes that we take toward the preservation of the environment during a period of explosive economic growth.

Whatever one might think about the data about global warming, nobody dismisses the concern of global climate <u>change</u> as something that's merely science fiction. It's plausible -- we've seen pictures taken from space not only of this country but of entire regions of the world -- and they look different than they did 10 years ago. Deforestation is a fact, not a theory. The limitations on the water supply are a major constraint to growth, not just some possibility. And the list goes on and on. We are increasingly faced with bumping up to the limits of what nature is willing to give us.

No nation facing these questions in an honest and democratic fashion can turn their head aside because none of us -- whatever business or industry we're in -- want to be in a situation where we can't take our kids or grandkids out in a natural environment and let them experience that for themselves and make their own choices.

Don't you see how clean coal technology is right in the middle of that?

It's not just this government and this Adminstration that ran on a politic platform of improving economic growth and creating jobs. That is the same platform that politicians now throughout the world are running on. They put their test of whether they're going to be elected or reelected ultimately on that economic growth, growing economic opportunities for growing populations. Make no doubt about it.

That requires the basic infrastructures of our countries -- power, electricity, transportation, water supply, legal and property rights -- to be in place. Without those foundations, no nation has ever had sustained economic growth. There are nations that have literally come and gone -- that's what archaeology is all about. You don't think there are going to be archaeologists in 500 years? What countries are they going to dig up? They'll dig up the ones that didn't sustain economic growth.

Now Jack and others are right in saying that our most abundant resource is coal. And you know the squeeze and the dilemna about the alternatives. I don't need to talk to this group about the turmoil and the economics encountered in the nuclear industry. There's not a serious, thoughtful thinker that can say that coal is not a part of the power future of this country. We know that. And we in this Administration are committed to seeing that the coal technologies of this country advance in a way that's compatible with the other interests that I outlined. The fact that right now we've run into the limits of nature and we're trying to figure out as a people what to do about that -- not just in our country but in others.

The government that the people in this room have been paying for through their tax dollars has made an enormous investment. We've done what many people are challenged to do; we've put our money where our mouth is through the Clean Coal Technology program as have many of our industrial partners. We have a number of projects and we have results. Some of those aren't what we expected them to be, but a lot are or are better. There is a track record.

The question that I have in my mind is this: will the industry and industry groups represented in this room, starting with the utility industry, be willing to step out and get ahead of the curve? Get ahead of the economic trend that they see coming? Or will they wait to be pushed along? And if they wait, will the trend overpower them and pass them by?

Look at the way that large industrial enterprises -- including utilities -- have evolved over the last 100 years. You know, it hasn't been that long since the advent of the corporation, the international corporation and the form of doing business where many people pool their capital and create large enterprises. If you can say anything about the history of the corporate enterprise, both in this country and abroad, it's that no company -- however big and perhaps even especially the big -- is immune to change.

And organizations which resist that change, the market overtakes. And it is overtaking them at an accelerated pace. You know, I don't come from the utility industry myself, and I've been told by people who are more famaliar with the industry than I what a conservative group this is -- made even more conservative by the fact that, in many cases, regulatory commissions have been able to use the benefit of hindsight to penalize without creating sufficient reward for risktaking.

But I'll tell you something. The most risky strategy for any industry, the utility included, is not to change, and not to try to remain in front of the trend. We can some day go look at the companies that make up the Dow Jones industrial average and look at who they were 30 years ago and who they are today. We can look at what people said about them 30 years ago and what people say about those same companies today.

You will see that the fastest growing companies, the companies that offer real security, are those who have put themselves at the forefront of technological change. Those that have missed the change in technology -- even by a mere 5 to 10 years -- are the ones who are struggling to survive. And they are surviving only by borrowing amounts of money they will not be able to repay unless they change their way of doing business and unless they change their technology.

We have a track record in the utilization of coal which reduces emissions and increases efficiencies in its use. Those who want to wait 5 or 10 years, to make sure that the rest of the industry goes before them, to take a wait-and-see attitute, see the safe thing as being behind the pack.

But that's not the safe place.

The safe place is to be right at the forefront of where that change is. We need to recognize that the trend of awareness of respect for our environment is one that is occurring worldwide, is one that transcends partisanship and ideology.

Yes, there are times when the regulatory commissions of the states and the federal government make mistakes. People in government make a lot of mistakes. I said earlier that some of the biggest companies were the ones finding themselves most in trouble in this world because they had become so successful that they resisted change, they resisted new technology. Well, the biggest enterprise of all is the government, and we've made plenty of mistakes.

Sometimes we within government -- I've only been here three months but have the identity association already -- those people in government, who look both to regulate industry and balance environmental concerns against concerns for growth, are struggling too. And we have vowed to do a better job and to take seriously what this week the President and Vice President will be preaching -- which is to view the taxpayers, the businesses, the employers of America as our customers, as people we must please and serve.

So by challenging you in the use of new technologies, I do not want to be presumptuous. I know that the knowledge and information that we have at the DOE and within government has only been purchased by use of other people's money -- the taxpayers' money. It is our obligation to get information into the hands of people as quickly as possible. I commit for all the employees of the DOE that we we will try to do that. If it means working long hours, if it means using the fax instead of a first class letter to assist you in the changing environment in which we're living, we will do that.

But ultimately, as we realize in this country and as other countries realize as well, government can only play a small part of the economy. It cannot run the economy; it cannot take most of the resources of the economy. It is going to be utilities and vendors who understand the regulatory framework with which they operate who are going to have to take some risks with these new technologies. We challenge you to do that.

There are many people who have helped in this program today. Jack Siegel has been a key player along with all the DOE employees who are here. I thank them for the work they've done in bringing you together as well as the Southern States Energy Board.

I think we will see conferences like this growing as time goes on and as people realize that power is not a matter of ideolgy or theology. When you read the facts, you will understand what electricity generation necessarily will be over the next two decades and that coal and clean coal technologies are squarely at the heart of that.

If anyone here in taking me up on my challenge -- whether a vendor, a utility, or a regulator -- takes a move that steps out in front and gets well ahead of the prospect of fines from the Clean Air Act and wants to set a new standard -- a standard that will endure for the year 2000 or 2010 -- and wants public recognition in support of taking that risk and implementing that new technology -- I encourage you to call us at DOE. It's part of our leadership role in this technological effort to highlight your efforts, to make sure that these efforts receive attention, and to make sure the message gets to the consumers of power who often take power for granted and only become aware of problems and take for granted the people who find solutions day in and day out.

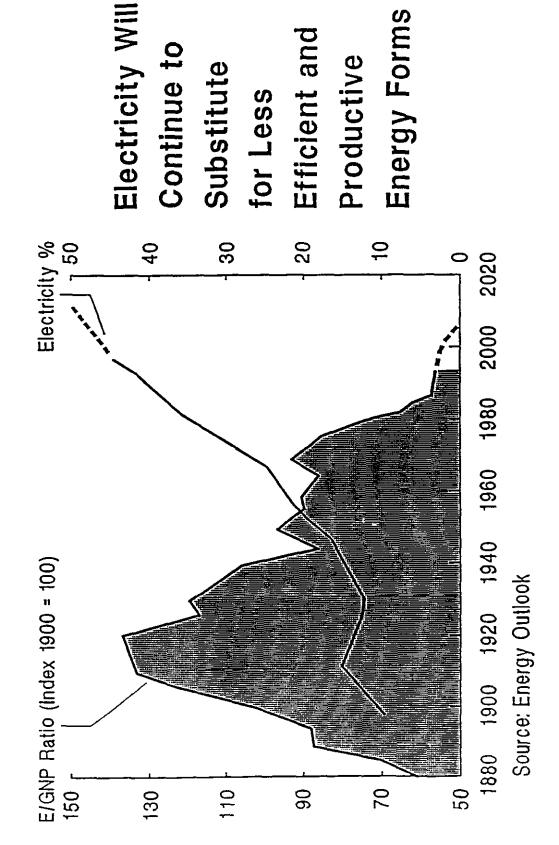
We will do what we can to express the support and appreciation of the poeple of the United States of America. We are as close as your telephone. We want to be accessible and we thank you for joining us this morning.

Plenary Session 1

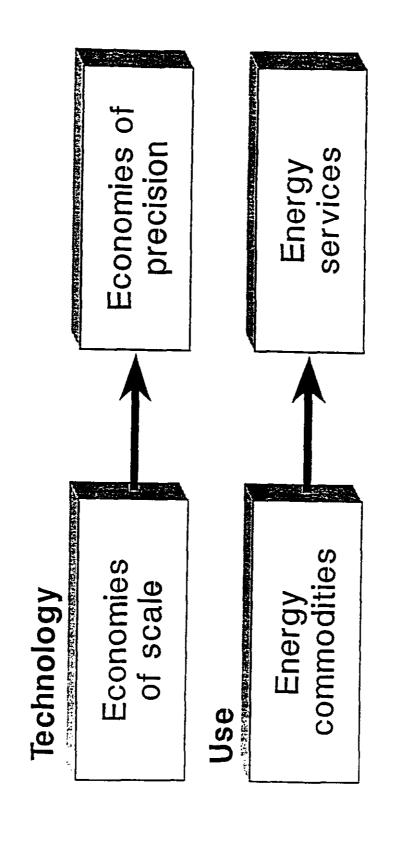


Kurt E. Yeager Senior Vice President Electric Power Research Institute Second Annual Clean Coal Technology Conference Atlanta, Georgia September 7-9, 1993

Energy Consumption/GNP vs Electricity Use

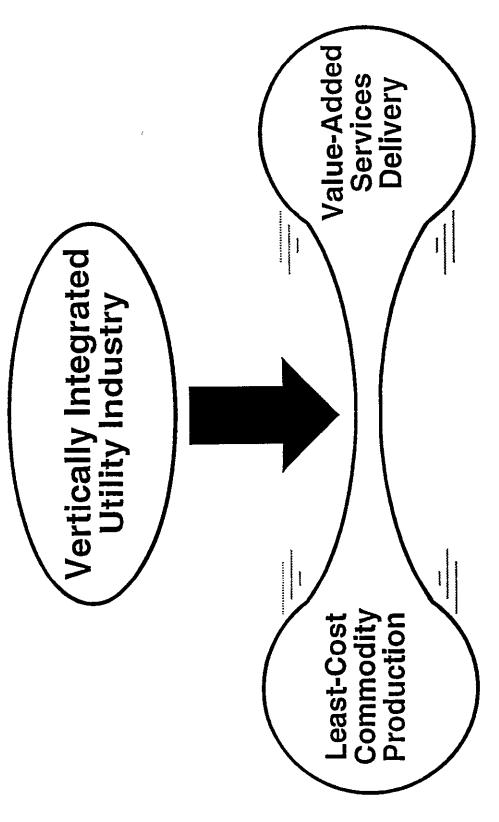


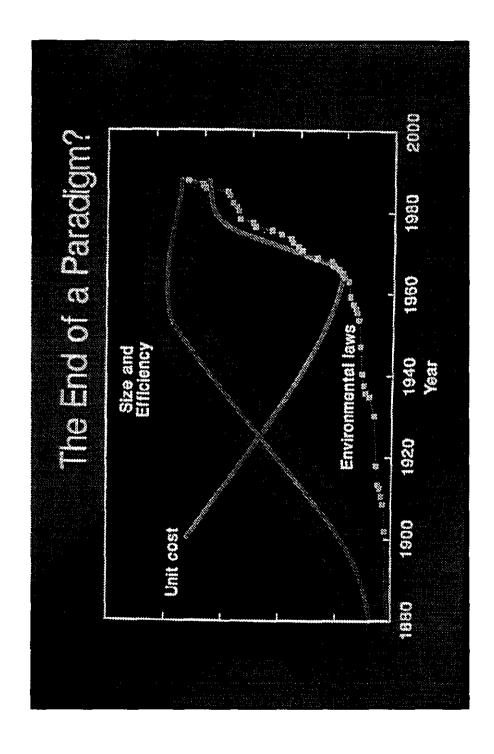
CHANGE DRIVERS

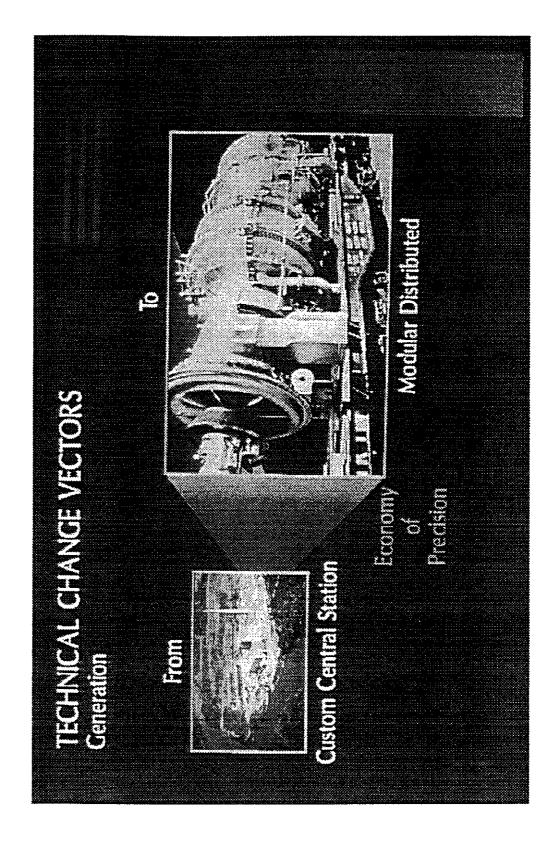


Derived from PG&E, 1991

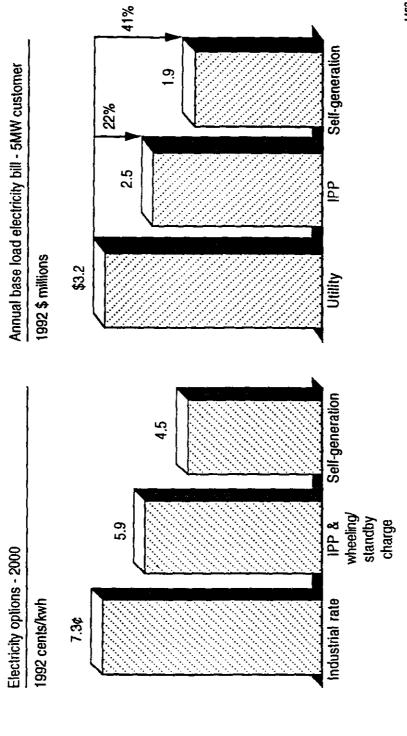
EVOLVING INDUSTRY STRUCTURE





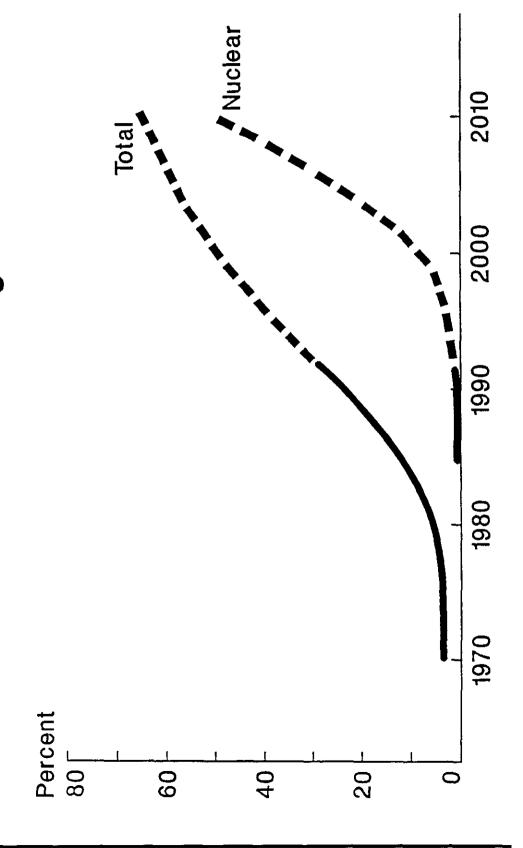


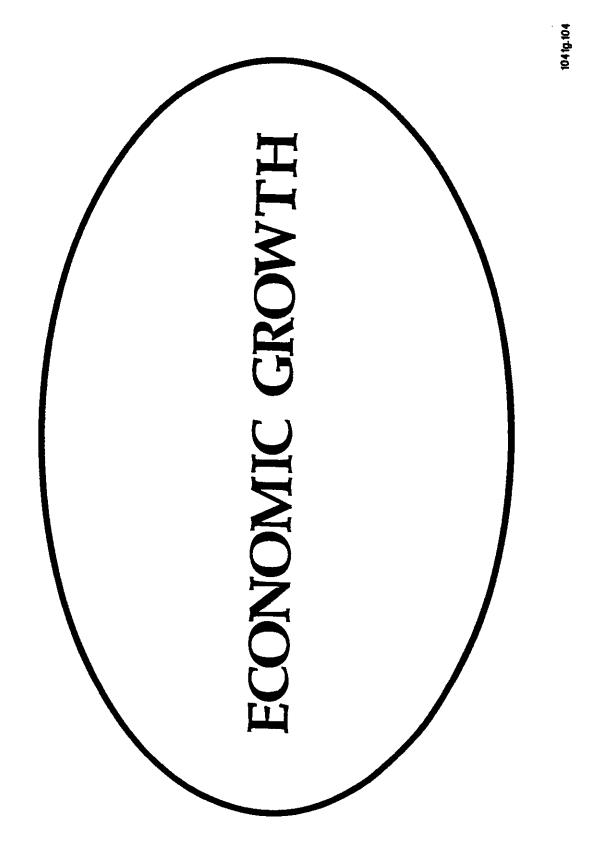
Customer Incentives Are Significant



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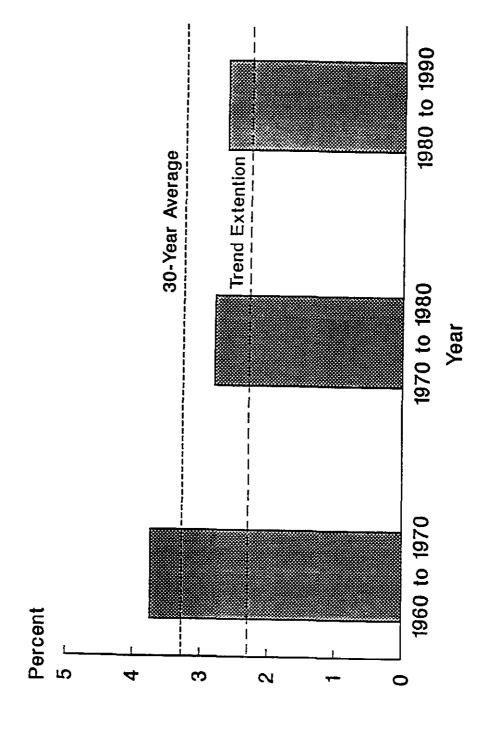
PERCENT OF GENERATING CAPACITY ≥30 Years of Age



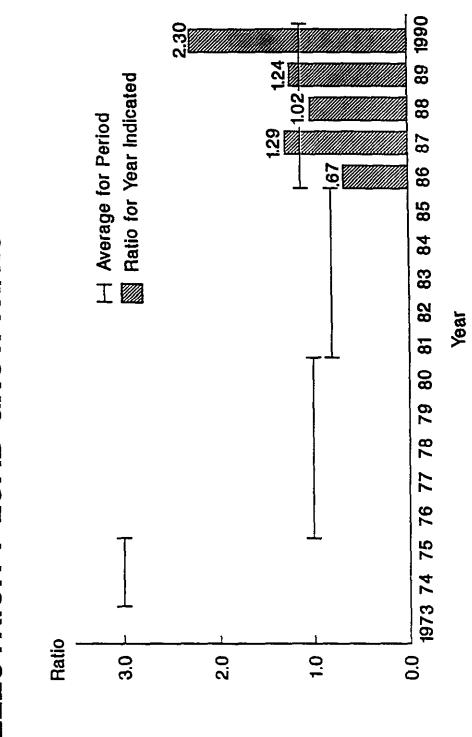


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ANNUAL GNP GROWTH RATES

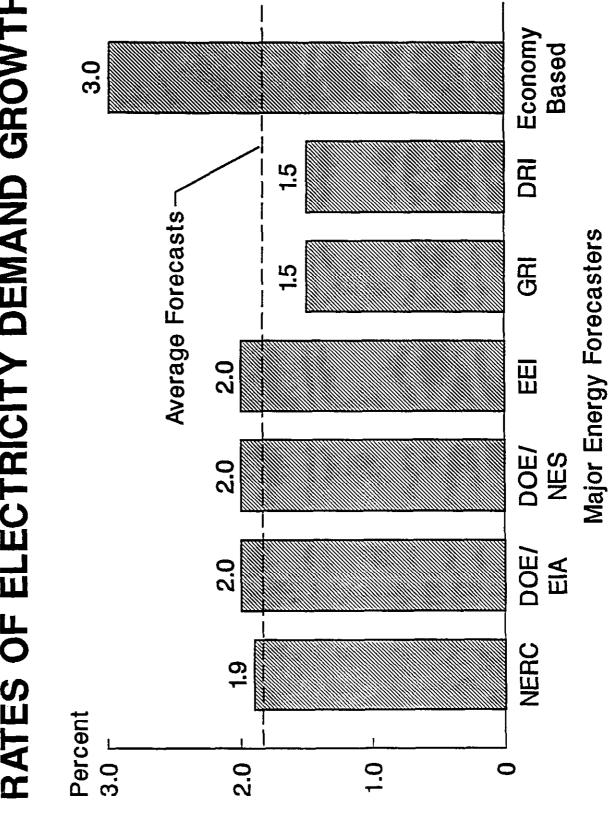


ELECTRICITY LOAD GROWTH/ECONOMIC GROWTH

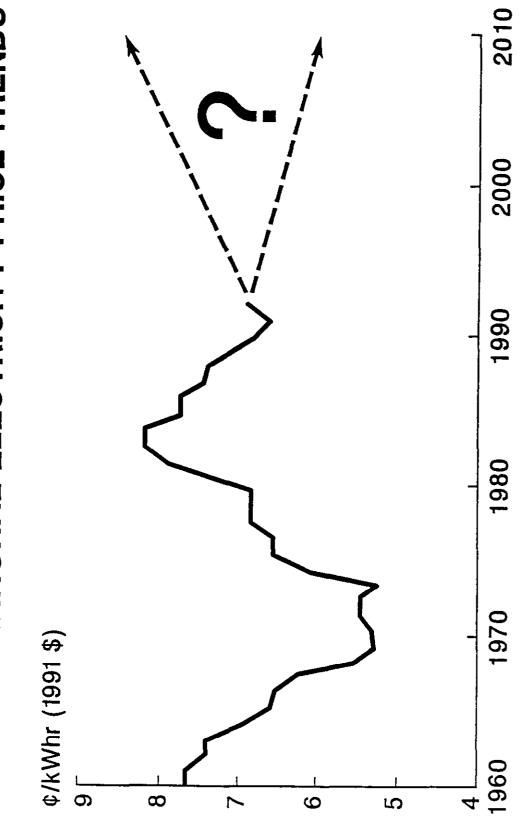


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RATES OF ELECTRICITY DEMAND GROWTH

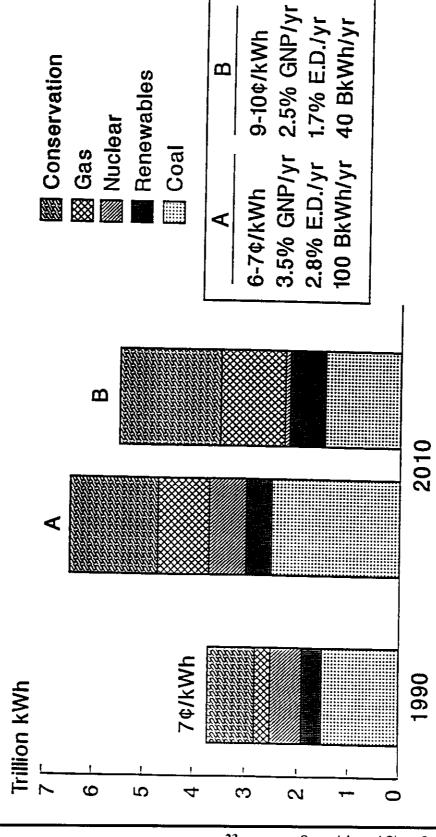


AVERAGE NATIONAL ELECTRICITY PRICE TRENDS

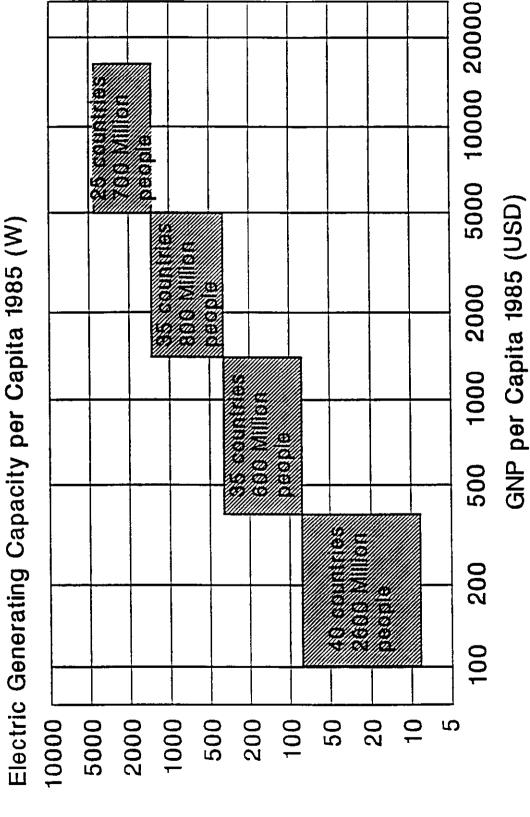


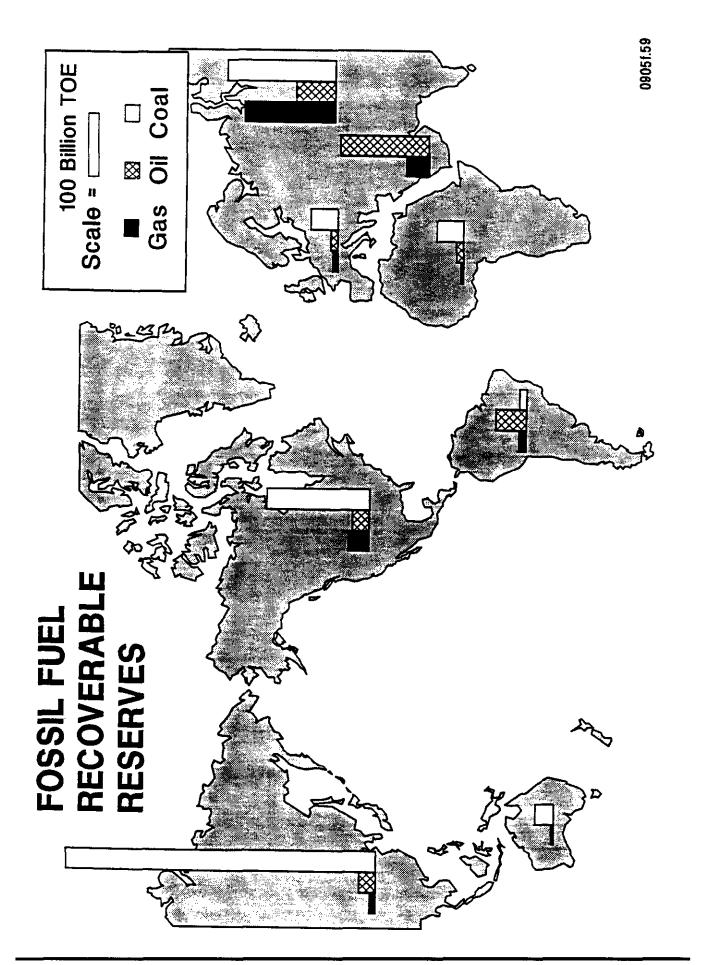
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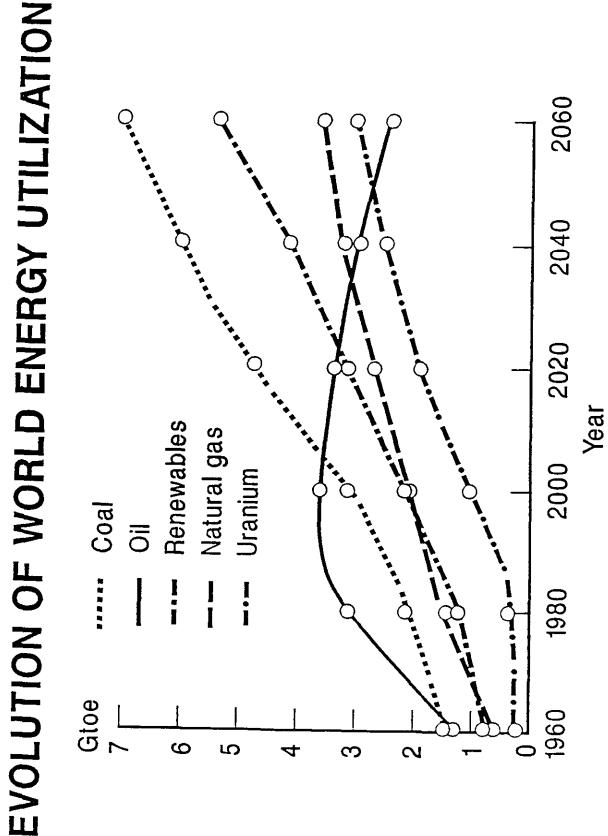
ALTERNATIVE FUTURES FOR ELECTRICITY GENERATION



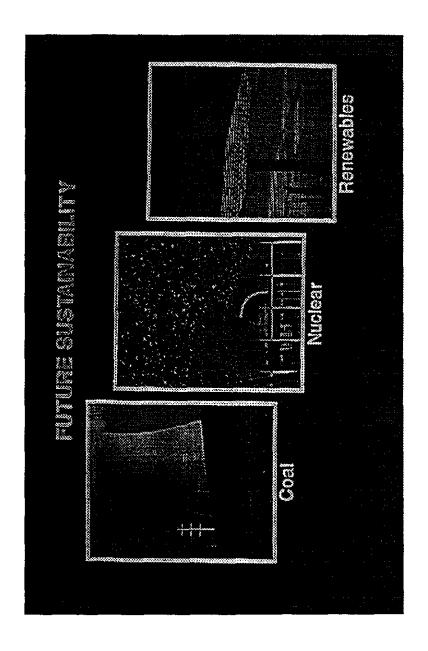
PROSPERITY AND ELECTRIC ENERGY







World Energy Conference - Conservation Commission, 1986





DISCLAIMER: The opinions and views expressed in this presentation are those of the author and do not necessarily represent the views of the Illinois Commerce Commission or other Commissioners.

Regulatory Climate for Clean Coal Technology into the Next Century

Remarks of Commissioner Lynn Shishido-Topel 2nd Annual Clean Coal Technology Conference Atlanta, Georgia September 8, 1993

I. Introduction

Good morning. I have been asked to talk about the regulatory climate for clean coal technology (CCT) into the next century. By clean coal technology, I mean new technology that uses coal more efficiently or cleaner in the combustion process than conventional techniques. The ability to use a domestically abundant fuel to meet increasingly stringent environment standards efficiently is certainly a valuable option to pursue. Rate of return regulation, with its capped authorized return and infamous disallowances is often criticized as a hostile environment for such promising but relatively risky investments. However, looking to the future, I think the most important issue for CCT is how well it will fare in a more competitive electricity generation industry with the kind of regulation such an industry implies. The next century is only seven years away, but many observers are predicting sea changes within the next five years. Will there be retail wheeling? To what extent? Will generation essentially become deregulated?

The focus of my talk today will be on how increasing competitiveness in the electricity generating industry may affect the regulatory climate for CCT generally. In this regard, I have two observations:

- The regulatory climate in the future may be more conducive to capital-intensive innovative technologies.

 However, CCT will have to develop faster payback times to do well in a more competitive future; and
- 2) that two things that could help it move in this direction are:
 - a. greater emphasis of government funding at the idea stage rather than at the commercial development stage; and
 - b. the careful use of incentives to achieve an efficient allocation of risk to utilities.

I also want to underscore the fact that state regulation is only one part of the picture. An increase in certainty over compliance standards for air toxics, co2, and nox, is also key to the future of CCT.

Let me start with a little background. State regulation is a creature of state statute. Therefore regulators do not have total discretion to craft regulatory devices or mechanisms. Fuel adjustment clauses, for example, had to be specially legislated in order not to run a afoul of legal restrictions against single issue ratemaking. Similarly, incentive regulation would require specific legislative authority and is not permitted currently by many state statutes.

An increasingly popular regulatory structure mandated by state statute is least cost planning, also known as integrated resource planning. The National Association of Regulatory Utility Commissioners (NARUC) defines IRP as "a way of analyzing growth and operation of utilities that considers a wide variety of both supply and demand factors so the optimal way of providing electric service to the public can be determined." The planning horizon is set out by state statute. In Illinois it is 20 years. Generally, the present value of revenue requirements of various options are compared. Long-lived, capital intensive projects with big upfront costs, and payoffs far into the future fair less well than projects with a lower upfront costs and faster payoffs.

An increasingly relevant question is how regulation will have to change to accommodate the changing environment inhabited by ratepayers and utilities. In this regard, one aspect of least cost planning process that may need to be considered is the planning horizon length over which various options are evaluated. As the generation industry becomes more competitive, it may become increasingly difficult to know with any degree of confidence what conditions will be in place 10 years from now, let alone 20. Will there be retail wheeling? What sort of technologies will competitors utilize? How will technological innovations spawned by a more competitive environment affect long-range planning assumptions? If planning horizons do shorten in response to a more uncertain, competitive environment, long-lived, capital intensive projects with payoffs far into the

future will have greater difficulty passing least cost screens. This is the major challenge I see for CCT.

Currently, it appears that the payoffs to CCT occur very far into the future and are not sufficiently large relative to the A recent PUF article described two instances in upfront costs. which, but for government grants, a cct project would not have been approved by state regulators. In addition, for one of the projects, even with the DOE funding, it was expected it would be 17 years before ratepayers saw benefits to the use of CCT. have no personal knowledge of the particulars of the cases aside from those reported in the article. However, these examples indicate that if the planning horizon under regulation is shortened, the amount of subsidies required to obtain acceptance of the cct project, all else constant, would have to increase. Now, state regulators are always happy to be offered federal funds to defray our costs. However, if the goal of cct research is to develop the most efficient and salable technology possible, increasing government subsidies in order to sustain otherwise uneconomic projects is unlikely to achieve this goal. Nor is this method likely to be practical. The two projects described in the article received 189 and 120 million dollars in federal grants, respectively.

This is not to say that there is no role for government subsidies. The classic problem for innovation is that because one cannot be compensated for all the benefits attributable to one's efforts there will be less than the optimal amount of

investment into innovation. Thus, potentially socially beneficial effort into technological innovation is often governmentally subsidized. Under this theory, however, subsidies should be applied where the ratio of private gain to social gain is smallest. The concept and initial pilot stage would appear to have smaller ratios than the commercial demonstration stage in which the utility participates. This is because while good ideas can take lots of effort to generate, you can't patent them. Furthermore, at the concept or pilot stage, much of the activity consists of understanding what are not good ideas and what won't work and no one will pay you very much for that, although it is valuable to have been done. At the commercial demonstration stage, however, the ideas generally have been proven and the benefit of a marketable technology can be made proprietary. In one of the cases I mentioned above, for example, the utility would have the right to profits from commercialization of the technology by other utilities.

It seems to me that constant innovation is going to be the name of the game so that a lot of attention should be paid to generating new ideas, techniques to reduce payback times and or reduce costs. It would therefore seem that the most important use of scarce government funding would be to help generate ideas rather than to assist commercial demonstrations. While some government subsidies may still be necessary, there should be less emphasis on government funding and more emphasis on entrepreneurial initiative at the commercial demonstration stage

so that the most promising technologies to commercially develop well be ferreted out. In this regard, traditional rate of return regulation has been criticized as providing little incentive for utilities to invest in uncertain technology and to operate As a result, it is argued that the current efficiently. regulatory climate is not conducive to innovative, entrepreneurial, activity. This view is based on the fact that the incentive to engage in innovative behavior is dependent on the expected gain and the risk of doing so. Under rate of return regulation, it is argued, the expected gain is insufficient to compensate for the anticipated risks. Under rate of return regulation, the utility is given the opportunity to earn a set authorized rate of return determined to be appropriate through formal hearings. Rates are a function of just and reasonable expenses and the return on the utility's approved rate base. Unreasonable and imprudent expenses or capital expenditures are disallowed. Untried technologies present a greater risk of disallowances due to construction cost overruns, management mistakes due to lack of experience with the technology, abandoned plant due to failed technology. Thus, it is argued that since great performance is not rewarded and bad performance is punished, there is no incentive for the utility to take risks that could be avoided by using more traditional technology. is also argued that there is little gain to cost-reducing investments since these gains would be eliminated at each rate In addition, since reasonable costs are passed through,

and because regulators cannot detect with certainty all unreasonable costs, the incentive to minimize costs is reduced. This characterization is not totally correct. Disallowances are tempered by a regulator's statutory concern with a utility's financial viability. In addition, due to regulatory lag, utilities can benefit between rate cases from cost-reducing activities. The timing of rate cases is largely up to the utility. However, disincentives may exist for relatively long payback, capital-intensive investments such as cct.

The view that the use of incentives could improve utility performance to the benefit of ratepayers is certainly not new. The debate has centered on how to apply them. The concern is that incentives would still be applied under a regulated structure and be subject to potential abuse. There is wide agreement that if not applied carefully, you can get perverse results.

However, as the industry environment changes, there is increasing agreement that regulation may need to change with it. One point of view is that incentives mechanisms are necessary to get utilities to develop and use skills similar to firms it will be competing with. Implicit in this view is that a greater entrepreneurial spirit may better able utilities to meet the increasingly costly and complex challenges of providing electricity in a cost-effective manner.

In this regard, the use of incentives whereby a utility is rewarded for superior performance in return for accepting certain

risks could increase the willingness of utilities to adopt CCT. In addition, it could allow entrepreneurial forces to reveal the more salable and beneficial technologies. Finally, if utilities are, say, willing to absorb cost overruns in exchange for the ability to profit from "underruns" relative to a benchmark incentive scheme, the upfront costs a utility will require from ratepayers to fund CCT investments should be lower than under rate of return regulation. However, it should be noted that these are general consequences that apply to other technologies as well. Thus, while conducive to CCT, the use of incentives alone will not necessarily assure its success.

In any case, I think that resolving the uncertainty over environmental rules on air toxics, Co2, and NOx is also key to the future of cct. Given the large investment required for CCT, great uncertainty over how future rules will affect the need to incur additional costs will influence the value of your investment could easily discourage such investment. Some observers credit this uncertainty for the relative lack of interest in cct for phase one compliance. This observation is NOSE. Supported by a recent research paper which finds that uncertainty over federal regulatory change after 3 Mile Island was more important than technological uncertainty in the decision to cancel or not invest in a nuclear plant.

II. Conclusion

In conclusion, CCT is of interest because it has the potential to use a very abundant fuel to meet environmental standards more efficiently than other means. Given the changing economic environment in the electric industry, CCT therefore should be viewed as a potential competitive strategy as well as a potential compliance strategy. The success of CCT will therefore depend in large part on how well CCT and the way it is developed will be able to adapt to the changing economic environment.



Potential air toxics regulations create a similar dilemma - emission regulations that out-distance economical, acceptable-risk technologies to address such regulations. Does the Clean Coal Program even address air toxics? There are significant problems in even determining the low concentrations of some of these elements in the coal or flue gas. For example, cadmium, selenium, and especially mercury are extremely difficult to measure.

Solid waste management. Will solid waste disposal regulations continue to get tougher? Can we find more ways to utilize these materials? Solid waste management or by-product utilization has become a major R&D priority for CONSOL. The question is: Should it become a higher priority for the Clean Coal Program?

Carbon dioxide emissions. Will we see CO₂ emission reduction regulations in the near future? If so, will the advanced power generation technologies be successfully demonstrated and ready to go at economics that make new or repowered coal-fired plants viable? Will hybrid technologies of gasification and fluidized bed combustion be possible long term solutions? Will advanced combustion technologies, like those being developed under the DOE Combustion 2000 Program, achieve thermal efficiencies of 50% or above? Many advanced technologies

won't be commercially viable until well after the year 2005. Will governments wait until then before legislating global climate change laws? Will the U.S. government facilitate the transfer of advanced combustion technologies to the developing countries? Can the world economy even afford global climate change emission reductions?

Now, let me quit beating the environmental regulators and address two other concerns. First, the deployment schedule of clean coal technologies. The ultimate success of the DOE Clean Coal Technology Program will be measured by the contribution that the technologies make to the environmental, economic, and energy future of our nation. Will utilities take the risks to deploy clean coal technologies? Will Public Utility Commissions give incentives to the utilities to take the needed risks? I hope the panel session Thursday morning on clean coal technology deployment and technology transfer addresses these concerns.

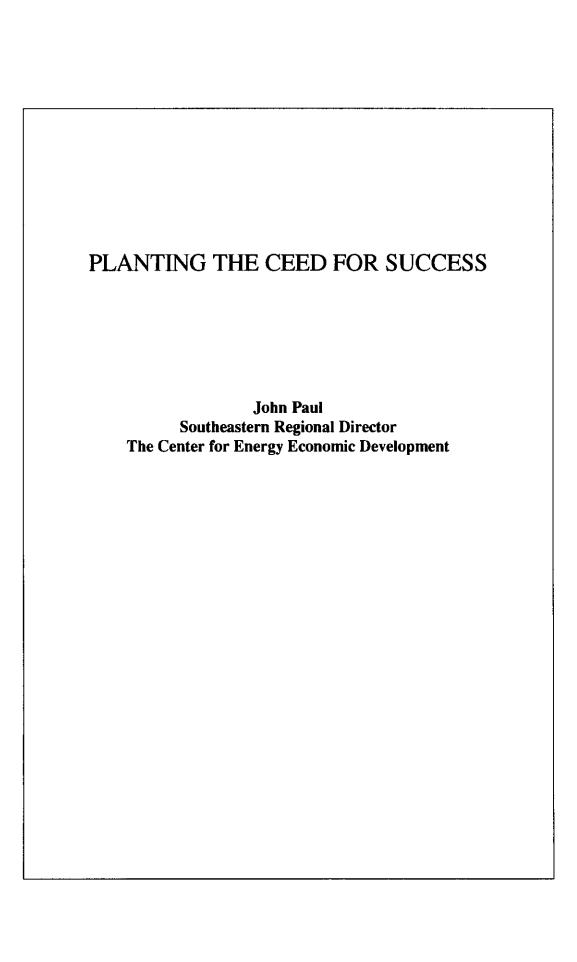
My final concern deals with energy policy and the definition of a clean coal technology. There have been initiatives to persuade PUCs to endorse co-firing of natural gas with coal by electric utilities as a so-called clean coal technology. Co-firing is fundamentally an unsound utility SO, control compliance strategy due to poor economics. Studies

have shown that scrubbing, coal blending, or even switching to lower-sulfur coals is economically superior to natural gas co-firing. Co-firing can be shown as an economic compliance strategy only when using unrealistically low gas prices that do not reflect the risks associated with volatile future gas prices. I believe natural gas best fits as an important resource for high-value applications such as home heating and transportation.

Even though we have such concerns about clean coal technologies, I want to conclude on a positive note. CONSOL remains committed to the commercialization of clean coal technologies. We want to applaud the many utilities across the nation that have and will take the risks to demonstrate and deploy these promising technologies. We applaud the state public utility commissions which have allowed utilities to take the economic risks to test these technologies. I also want to thank Senator Byrd of West Virginia for his strong support of the Clean Coal Program, especially when it came time for budget appropriations. As we all know, coal is our most important long-term natural energy resource. Clean coal technologies can help to use it efficiently, economically, and in an environmentally acceptable manner.

I think we are going to find that the Clean Coal Program has achieved remarkable results—creating legitimate options for emissions control. It will be a major disappointment to me if we cannot celebrate those successes or applaud DOE, the coal industry, and others for spending large amounts of money merely because political groups with short attention spans, have shifted their attention to the new "politically correct" issues even before the current one is solved. Though it is frustrating to solve problems, and concurrently find that public interest has moved on, we should feel proud of our accomplishments in developing clean coal technology.

Thank you.



In 1944, over half of the American people heated their homes with coal. Even in the mid-to-late 1940s, coal was the favored heating source for most people relative to gas and oil. Coal was part of peoples' everyday lives. Sure, there were the negatives to using coal as with any fuel - people then and now think of air pollution scenes of Pittsburgh in the '40s -- but people also knew, personally, the benefits of coal.

Well, since the '40s, new technologies have cleaned the air in Pittsburgh. But coal as a home heating source has become nearly extinct -- and by extension, coal's familiar benefits have disappeared from view for most people.

In this generation, coal as an energy source has become something of an abstraction -- in many ways like nuclear power. People never see coal except when there are problems; Tragic but thankfully infrequent mine accidents, transportation problems resulting from a disruption or derailment, and blame for a range of environmental problems.

Today, coal's benefits are largely invisible. But coal's problems are very visible and easy targets for the media. And as coal has become more of an abstraction, false information and negative images brought by anti-coal forces -- which include the media, environmental groups and competing fuels -- have become more easily imbedded in the public mind and are trending more negative. Why? Because there are precious few countervailing positive images of coal -- we no longer have the personal experience, as with home heating, or we fail to recognize that the electricity that runs the conveniences in our daily living is, in fact, the modern manifestation of coal.

This model of "what's gone wrong," actually poses two challenges:

First, the issue of coal's negative image. There are negative imbedded attitudes about coal and coal use across broad cross-sections of the American people, and targeted audiences of political/social activists. These negative attitudes are trending worse and are driven by organized opponents, and political agendas.

Secondly, there is the hurdle of doing something about it. The relative good times for the coal industry -- we have almost doubled our production in the last 20

years -- have masked the serious slippage in public opinion. Production is up. Public opinion is down. This phenomena sets the industry up for a fall, if left unattended. An evading image is a ticking bomb.

So, what to do? It is very clear we have a major challenge on our hands. The facts I have just recited, and a series of delayed or cancelled coal units throughout the U.S., particularly a number of proposed coal fired independent power projects in the South, caused CEOs from several major railroads to review options aimed at addressing the problem. Those discussions led to interaction with their counterparts at major coal companies, and a new movement was formed. Major resources were joined to deal with a major challenge.

For those of you familiar with the history of the rail and coal industries you know that the sharing of resources under this new common banner - The Center for Energy and Economic Development - or CEED, is no small miracle. For those of you not familiar with the history of these two major industries let me simply say we have had a very torturous and often openly hostile relationship. Fortunately, rail and coal leadership recognized the overall, long-term good of both entities required the subordination of parochial interests and conflicting positions on specific national issues.

CEED has been organized to advocate responsible energy policy - a policy that does not discriminate against coal. Where there is coal - there is low-cost electricity and economic development. It is an umbrella under which a broad coalition of business and individual interests can cooperate.

The CEED process began with a comprehensive public opinion research program that would allow us to understand attitudes and opinions about energy and economic development, more specifically coal, and related issues. We reviewed the public opinion history of coal beginning with the first national survey in 1944, and then is December of 1992 we held a series of qualitative focus discussions in Tampa, Hartford, Denver and Indianapolis. In each city there was a discussion between business leaders and environmental activists, and one with the general public. In January of 1993, the focus groups were followed by a quantitative assessment of national opinion measuring trends, and collecting demographic and geographic differences.

Let me share a few observations that resulted from the focus groups and survey:

Slide Public Perception of Fuel Used to Generate Electricity in the U.S.

Slide Public Vision of Future Fuel Use to Generate Electricity in 10 Years

Slide	Public Knowledge About Coal - Electricity
	Coal Provides More Than 25%
Slide	Public Perception vs Reality of Coal Use
Slide	How Likely We Will Run Out of Coal in 50 Years
Slide	New Coal Plant in Your Area
Slide	Acceptability of High Technology Plant In
	Your Area
Slide	Future Importance of Coal

These survey results and the CEO level discussions led to the establishment of a plan of action -- that action was the creation of CEED.

Slide CEED

There are numerous industries and individuals economically allied with the coal industry and share concerns about coal's image. Where there are shared concerns there should be shared resources. These shared resources will be organized to produce positive education and outreach programs to business, the media and policy makers. CEED has been established to fulfill that mission. CEED will produce and sustain a long-term education and information effort to communicate messages about coal, the U.S. economy, new technologies and environmental progress and compatibility.

Slide Regional Organizations

We are a single purpose organization created for the purpose of keeping the coal option a viable alternative for utilities, IPPs and industrial users, funded by eight of the Class I railroads and coal companies representing more than 50% of the total U.S. production. The membership recognizes that it took us a long time to get into the predicament we find ourselves in and there will not be an overnight solution; therefore, there is a long-term commitment to the program.

the individual situation.

D.C. Administrative Office
Regional Offices
Member services and facilities/small staff
Not a beltway institution
Not a typical issue organization or coalition
Board of Directors/Regional and State Steering
Committees
Deal with each area or state in a manner that fits

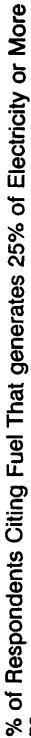
Slide Immediate Goals
Slide Communication Tools

Basically, we plan to "get out among them", "show up", and generally establish a presence to insure that the real facts get out. We intend to build out a single, straightforward, reality-based program that builds off the unambiguous strengths of coal, while recognizing existing public perceptions. This is not an anti-other fuels program, but we do want a level playing field where coal is part of the business decision process. This will be accomplished through a true grassroots effort that will energize individuals and entities, and in the long term establish coal's image as a fuel of the future -- a high technology product and a critical American asset that touches the lives of most Americans.

Thank you for your attention. I will be happy to answer questions or provide you with information on how you can join with us in this most important effort.



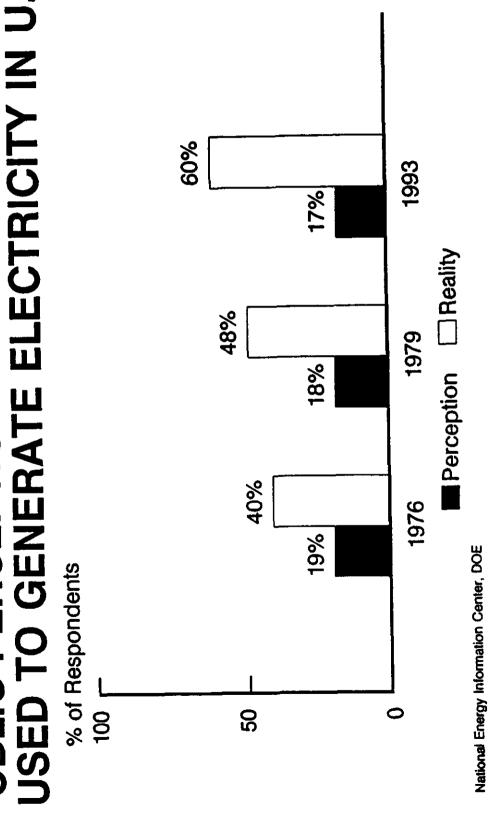
PUBLIC PERCEPTION OF FUEL USED TO GENERATE ELECTRICITY IN U.S.





79 ABC News/Harris & Associates Polf '83 Whithlin for the Center for Energy and Economic Development

PUBLIC PERCEPTION VS. REALITY OF COAL USED TO GENERATE ELECTRICITY IN U.S.

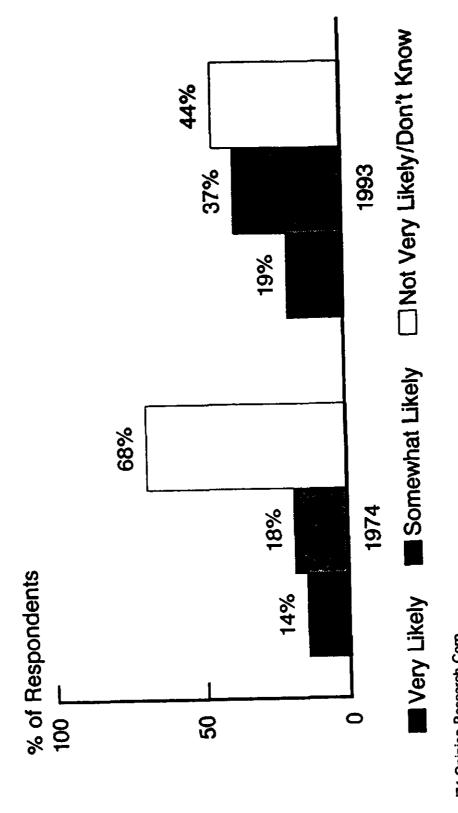


79 Harris & Associates 93 Wirthlin for the Center for Energy and Economic Development

76 Harris & Associates

- 60 -

HOW LIKELY IS IT THAT THE U.S. COAL SUPPLY WILL RUN OUT IN 50 YEARS?



74 Opinion Research Corp.

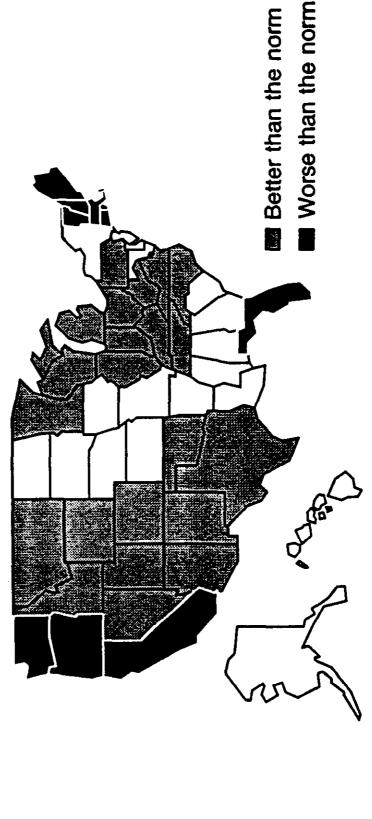
'93 Whithlin for the Center for Energy and Economic Development

NEW COAL PLANT IN YOUR AREA? 1993 WVA/PA/KY % 35% 57% ■ Not Sure % 1993 FLA 52% Not Acceptable 41% 10% 1993 Nat'l 43% 47% ☐ Acceptable **%** 1991 Nat'i 51% %09 %

'91 Yakelovich Clancy Schulman '93 Wirthlin for the Center for Energy and Economic Development

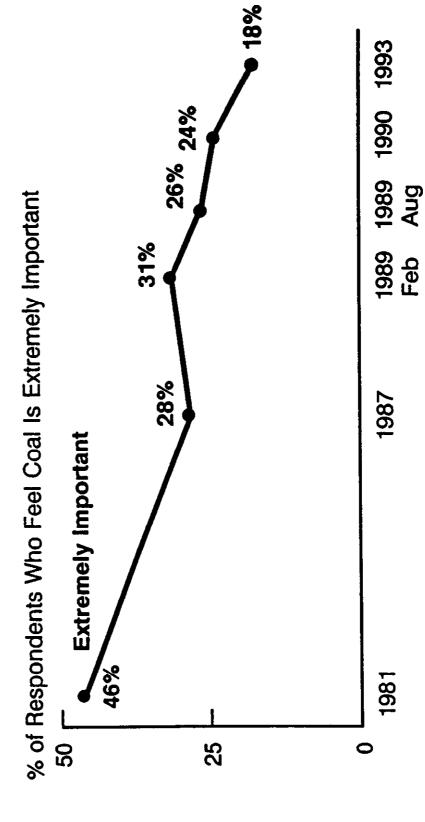
"Acceptability of Building High Technology Coal Plant in Your Area" PUBLIC ATTITUDE ABOUT COAL

National Norm: 47% of respondents approve; 43% of respondents disapprove



'93 Wirthlin for the Center for Energy and Economic Development

OUR FUTURE ENERGY NEEDS: THE IMPORTANCE OF COAL



'81,'87,'89,'90 Cambridge Report Poll for U.S. Council for Energy Awareness '93 Wirthlin for the Center for Energy and Economic Development



Northern Southern Midwestern CEED REGIONS Western

AND ECONOMIC DEVELOPMENT THE CENTER FOR ENERGY Immediate Program Goals

Organize and muster a broad base of pro-coal interests in states and communities:

- Plan and produce education and communications programs
- Neutralize efforts aimed at restricting the use of coal
- Reach out to decisionmakers in government, business, education, and the media
- Produce local pro-active programs in support of new high-technology coal/electric power plants
- Place expert witnesses at state and local regulatory and legislative proceedings, externality and siting hearings

AND ECONOMIC DEVELOPMENT THE CENTER FOR ENERGY **Communications Tools**

Recruiting and advocacy videos

Ceed public affairs kit

Electronic interactive information network

Luncheon

Speaker introduced by:

Jack S. Siegel,

Acting Assistant Secretary for Fossil Energy,

U.S. Department of Energy

nd Annual Clean Coal Te	chnology Conference	e - 70 -	 	

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CCT CONFERENCE ADDRESS
MICHAEL K. REILLY
SEPTEMBER 7, 1993 -- FINAL

THANK YOU LADIES AND GENTLEMEN.

ON BEHALF OF AMERICA'S COAL INDUSTRY, I WANT TO COMMEND THE DEPARTMENT OF ENERGY AND THE SOUTHERN STATES ENERGY BOARD FOR PUTTING ON THIS SECOND ANNUAL CLEAN COAL TECHNOLOGY CONFERENCE.

IT IS THROUGH GATHERINGS SUCH AS THIS THAT FACTS AND FINDINGS ARE ACCUMULATED AND ASSESSED. FROM A SIFTING AND SORTING OF THE FACTS, INDIVIDUAL DECISIONS EMERGE. AS THE DECISIONS MOUNT, CONSENSUS FORMS. AND FROM CONSENSUS FLOW THE ACTIONS THAT TURN PROMISE INTO REALITY.

I AM HERE TODAY IN THREE CAPACITIES. FIRST, AS CHAIRMAN OF THE NATIONAL COAL ASSOCIATION, WHICH REPRESENTS THE COMPANIES THAT SUPPLY MOST OF AMERICA'S LARGEST DOMESTIC ENERGY SOURCE.

I AM ALSO HERE AS CHAIRMAN AND CHIEF EXECUTIVE OFFICER OF ZEIGLER COAL HOLDING COMPANY. THE ZEIGLER FAMILY OF COMPANIES PRODUCES MORE THAN 40 MILLION TONS A YEAR AND HOLDS RESERVES OF 3.5 BILLION TONS. ZEIGLER IS THE NATION'S LARGEST INDEPENDENT COMPANY DEVOTED SOLELY TO COAL.

FINALLY, I AM HERE TODAY AS A TANGIBLE SUPPORTER OF CLEAN COAL TECHNOLOGY THROUGH THE ENCOAL MILD-GASIFICATION PLANT... AT ENCOAL; LOCATED AT OUR BUCKSKIN MINE NEAR GILLETTE, WYOMING, WE ARE IN PARTNERSHIP WITH THE DEPARTMENT OF ENERGY THROUGH ITS EXCELLENT CLEAN COAL TECHNOLOGY PROGRAM.

ENCOAL USES LOW-RANK, SUB-BITUMINOUS COAL AND PRODUCES TWO HIGH-RANK FUELS OF LOW SULFUR CONTENT. THESE INCLUDE A LIQUID FUEL THAT SUBSTITUTES DIRECTLY FOR NO. 6 FUEL OIL... AND A SOLID PROCESS FUEL WITH A MUCH HIGHER HEATING CONTENT THAN THE FEEDSTOCK. WE SEE IN ENCOAL PROGRESS IN TWO ENVIRONMENTS--THE ECONOMIC AND THE NATURAL. THIS IS THE PROMISE OF TECHNOLOGY, AND IT IS ONE THAT IS BEING PLAYED OUT WITH VARYING DEGREES OF SUCCESS ACROSS THE UNITED STATES.

I WAS ASKED TO SPEAK TO YOU TODAY ON THE TOPIC OF...COAL: FUEL OF CHOICE AND FUEL OF NECESSITY. AND WHILE, ON ITS SURFACE, THE TOPIC MIGHT HAVE SOMETHING OF A GENERIC QUALITY TO IT, AT ITS ESSENCE IS THE CORE OF THE CHALLENGES THAT WE'VE FACED, AND THE OPPORTUNITIES THAT AWAIT US.

FOR THE STORY OF COAL IS THAT OF A LOVE-HATE RELATIONSHIP
THAT EXTENDS BACKWARDS MANY CENTURIES. COAL HAS SLOWLY...
QUIETLY... STEADILY CARRIED THE PROGRESS OF ENTIRE
CIVILIZATIONS UPON ITS BROAD SHOULDERS. YET WHILE COAL HAS
OFTEN BEEN THE FUEL OF CHOICE...IT HAS RARELY BEEN ACCEPTED
BY THE GENERAL PUBLIC WITHOUT RESERVATION. COAL HAS BEEN
VIEWED NOT JUST AS A FUEL OF NECESSITY BUT AS A NECESSARY
EVIL... SOMETHING TO GET US THROUGH UNTIL WE CAN FIND A TRULY
GOOD FUEL.

THIS HAS BEEN THE CASE, DESPITE THE FACT THAT OTHER ENERGY ALTERNATIVES HAVE CONSISTENTLY FAILED TO MATCH COAL'S STABILITY... CONSISTENTLY FAILED TO MATCH COAL'S AVAILABILITY... AND CONSISTENTLY FAILED TO MATCH COAL'S PRICE.

WHEN I CONSIDER THE CLAIMS OF COMPETING FUELS... AND IT MAKES NO DIFFERENCE WHETHER YOU ARE DISCUSSING NUCLEAR ENERGY IN THE NINETEEN-SIXTIES OR NATURAL GAS IN THE NINETIES... I AM REMINDED OF THE STORY ABOUT A LITTLE BOY WHO WALKED INTO A CANDY STORE AND ASKED FOR A PISTACHIO ICE CREAM CONE. WHEN THE OWNER TOLD HIM IT WOULD COST A DIME, THE BOY SAID THAT THE STORE ACROSS THE STREET ONLY CHARGED A NICKEL.

"SO WHY DON'T YOU BUY IT THERE?" ASKED THE OWNER.

"BECAUSE THEY ARE OUT OF PISTACHIO," REPLIED THE BOY.

"WELL," SAID THE OWNER, "IF I WAS OUT OF PISTACHIO, I'D CHARGE A NICKEL, TOO."

FUEL OF CHOICE. FUEL OF NECESSITY. BUT IF YOU LOOK TO COAL'S PLACE IN THE NATION TODAY, AND FROM WHERE IT HAS COME, THAT IS NOT REALLY THE PROPER ORDER. AFTER ALL... COAL'S ASTOUNDING ABILITIES TO PROVIDE ABUNDANT, INEXPENSIVE AND RELIABLE FUEL WAS RECOGNIZED BY THE CHINESE AS FAR BACK AS 300 A.D., AND IN WHAT IS NOW AMERICA BY THE 1400S.

THOSE WHO ARE NOW READY TO PROCLAIM THE DEATH OF COAL WOULD DO WELL TO REMEMBER THAT THE FIRST ENVIRONMENTAL PRONOUNCEMENT EXPECTED TO DOOM THE COAL INDUSTRY WASN'T THE CLEAN AIR ACT OF 1970, OR ITS AMENDMENTS IN 1990--BUT AN EDICT FROM ENGLAND'S KING EDWARD THE FIRST IN THE EARLY 1300S.

NO SOONER HAD FOURTEENTH CENTURY ENGLAND DISCOVERED THE WONDERS OF COAL THAN THE KING CAME OUT WITH A HARSH ATTACK AGAINST, QUOTE, THE STINK AND BADNESS OF THE AIR AND THE DESTRUCTION OF THE FRUIT TREES. THERE IS NO RECORD, INCIDENTALLY, THAT HE MADE ANY REFERENCE TO NITROUS OXIDE EMISSIONS OR CLIMATE CHANGE.

IF YOU'RE CYNICAL, YOU CAN DRAW FROM THIS STORY THAT COAL HAS ALWAYS BEEN UNPOPULAR...AND ALWAYS WILL BE.

BUT WHILE OUR CRITICS WOULD DWELL ON <u>ONLY</u> THE UNPOPULARITY OF COAL, THIS APPROACH CLEARLY MISSES THE POINT. FOR IF IT IS EVIDENT THAT THE CLASHES OF COAL WITH THE DEMANDS OF ENVIRONMENTALISTS HAVE A SEVEN-CENTURY LEGACY, IT IS ALSO QUITE CLEAR THAT COAL HAS NOT ONLY SURVIVED DURING THAT TIME... BUT IT HAS THRIVED.

ON AN EMOTIONAL LEVEL, COAL MAY NOT HAVE ALWAYS BEEN THE FUEL OF CHOICE. IT HAS BEEN, THOUGH, AND WILL CONTINUE TO BE, THE FUEL OF NECESSITY. WHAT I BELIEVE WE ARE CONCERNED WITH TODAY, AND EVERY DAY, IS EXPLORING NEW WAYS TO BRIDGE A NATION'S WANTS WITH A NATION'S NEEDS. WE WILL BE SATISFIED IF THE PUBLIC VIEWS COAL AS A NECESSITY. BUT I, FOR ONE, WOULD FEEL MUCH MORE SECURE IF THE PUBLIC VIEWS COAL THE WAY YOU AND I VIEW COAL... AS THE BEST SINGLE SOURCE OF ELECTRICITY IN THE COUNTRY TODAY.

WHY IS COAL THE FUEL NECESSITY? THE ANSWER LIES IN THE PRODUCT VIEWED BOTH SEPARATELY AND IN CONJUNCTION WITH THE OTHER AVAILABLE FUELS.

TO APPRECIATE THE BROAD SHOULDERS OF COAL IN OUR NATION'S LIFE, YOU DON'T NEED TO LOOK AT ITS MAJOR IMPORTANCE IN SETTLING THE WEST BY FUELING THE STEAM ENGINE. AND YOU NEEDN'T LOOK AT ITS CRUCIAL ROLE IN SERVING AS THE SPARK THAT IGNITED OUR INDUSTRIAL REVOLUTION. NO, TO VIEW OUR INDUSTRY'S BEHIND-THE-SCENES SUPPORT, YOU SIMPLY NEED TO PICK UP ANY NEWSPAPER OR MAGAZINE. THERE, HIDDEN IN A WORLD THAT HAS FOR TOO LONG TAKEN IT FOR GRANTED, ARE THE HUNDREDS OF STORIES THAT ILLUSTRATE HOW COAL AND ELECTRICITY CONTRIBUTE GREATLY TO OUR EVERYDAY LIVES.

IN PHILADELPHIA, COAL ASSISTED IN LASER SURGERY WHEN DOCTORS PERFORMED A BREATHTAKING OPERATION TO SEPARATE THE JOINED HEARTS OF SIAMESE TWINS. IN DENVER, COAL POWERED THE MICROPHONE AND VIDEO SCREENS THAT ENABLED THE POPE TO SPEAK TO HUNDREDS OF THOUSANDS OF HIS FOLLOWERS. AND HERE IN ATLANTA, COAL WILL BE A PARTNER... SILENT AS ALWAYS... IN ENERGIZING THE SPECTACULAR DISPLAY THAT WILL BE THE 1996 SUMMER OLYMPICS.

COAL LETS THE NEON LIGHTS SHINE BRIGHT ON BROADWAY. COAL FUELS THE COMPUTERS THAT COUNT DOWN THE SPACE SHUTTLE LAUNCHES. COAL ENERGIZES THE AUTOMAKER'S TOOLS, THE TEACHER'S CLASSROOMS AND THE BAKER'S OVENS FROM ALASKA TO FLORIDA.

COAL IS VAST AND ABUNDANT. IT CONSTITUTES 90 PERCENT OF THE NATION'S FOSSIL FUEL RESERVES--NEARLY 300 YEARS WORTH. AND IT ACCOUNTS FOR MORE OF THE NATION'S ELECTRICITY GENERATION THAN ALL OTHER FUELS COMBINED.

COAL'S LONGSTANDING USE CAN BE ATTRIBUTED TO SOME COMMON ADJECTIVES. CHEAP. ABUNDANT. DOMESTIC. RELIABLE.

IN A WORLD BLOATED WITH SLOGANS, THESE TERMS MAY SOUND HOLLOW. BUT OUR NATION HAS NEVER GONE TO WAR TO PROTECT OUR COAL INTERESTS, AS IT HAS WITH OIL. OUR NATION HAS NEVER SEEN ITS FACTORIES AND SCHOOLS CLOSE BECAUSE OF INSUFFICIENT COAL, AS THEY DID DURING THE NATURAL GAS SHORTAGES OF THE MID 1970S. AND OUR NATION HAS NEVER SEEN THE DRAMATIC PRICE VOLATILITY OF COAL THAT IT HAS SEEN WITH A VARIETY OF OTHER FOSSIL FUELS.

NONETHELESS, THERE ARE MANY PEOPLE WHO TODAY PREDICT COAL'S DEMISE, FEELING SURE THAT THE EFFECTS OF THE CLEAN AIR ACT WILL CRIPPLE THE INDUSTRY. TAKE THIS QUOTE, FOR INSTANCE:

"ALTHOUGH OUR INDUSTRY HAS MANY SERIOUS ECONOMIC AND REGULATORY PROBLEMS FACING IT TODAY, NONE ARE AS THREATENING AS THE CLEAN AIR ACT. THE SULFUR RESTRICTIONS IMPOSED BY THIS ACT ARE SEVERELY RESTRICTING THE ELECTRIC UTILITIES IN MEETING THE ENERGY DEMAND. THE LOW SULFUR FUELS REQUIRED BY THIS LEGISLATION ARE NOT GENERALLY AVAILABLE TO THE UTILITIES. SULFUR LIMITS HAVE BEEN SET WITHOUT REGARD FOR THE CURRENT STATE OF EMISSION CONTROL EQUIPMENT WHICH, DESPITE SIGNIFICANT PROGRESS, HAS NOT YET REACHED THE STATE OF A PROVEN FULL-SCALE COMMERCIAL PROCESS." END OF OUOTE.

THIS IS TYPICAL OF THE RHETORIC OF THE DAY, AND INDEED THERE ARE ISSUES THAT OFFER MAJOR SOURCES OF CONCERN. BUT WHAT IS NOTABLE ABOUT THIS STATEMENT IS THAT IT CAME FROM MY OWN COMPANY'S ANNUAL REPORT IN 1970, IN RESPONSE TO THE ORIGINAL PASSAGE OF THE CLEAN AIR ACT.

SINCE THAT TIME, ZEIGLER HAS INCREASED ITS SIZE BY A FACTOR OF MORE THAN 10 TIMES. BUT, MORE IMPORTANTLY, THE UNITED STATES COAL INDUSTRY HAS GROWN A HEALTHY 62 PERCENT. LET ME ASSURE YOU THAT WE AT ZEIGLER--AND MOST PEOPLE IN THE INDUSTRY--WOULD BE QUITE CONTENT WITH ANOTHER 20 YEARS AS "BAD" AS THE PAST 20.

REALITY AND PERCEPTION ARE NOT ALWAYS THE SAME, OF COURSE, AND COAL'S IMAGE HAS LONG LAGGED BEHIND ITS GRAND FUNCTION. AT THE SAME TIME, THE POWERS THAT BE LARGELY CONTINUE TO OVERLOOK COAL'S ACCOMPLISHMENTS AND PROMISE IN FAVOR OF THE POPULAR ENERGY OF THE DAY.

IN THE 1960S, <u>NUCLEAR</u> ENERGY WAS GOING TO BE THE FUEL TO MAKE ALL OTHER FUELS OBSOLETE. MORE RECENTLY, NATURAL GAS HAS ATTRACTED AN INSTITUTIONAL FOLLOWING IN WASHINGTON AND ELSEWHERE. AND RENEWABLE ENERGY CONTINUES TO CAPTURE THE IMAGINATION OF OUR NATION'S COUNTERCULTURE.

BUT TAKE A CLOSE LOOK AT ONE OF TODAY'S ENERGY DARLINGS...
NATURAL GAS... AND SOMETHING INTERESTING OCCURS. BECAUSE IF
YOU ATTEMPT TO ERASE COAL FROM THE AMERICAN SCENE AND
SUBSTITUTE NATURAL GAS, YOU ARE NOT LEFT WITH A WONDERFUL NEW
WORLD. IN FACT, YOU ARE LEFT WITH... A LOT OF QUESTION
MARKS.

THE SIREN SONGS OF INDUSTRIES LIKE NATURAL GAS ARE PLAYED LOUDLY TODAY. BUT EVIDENCE SHOWS THAT THEIR PROMISES OF RELIABILITY AND PRICE SIMPLY CANNOT BE BACKED UP.

THERE IS A REASON THAT NATIONALLY, SINCE 1970, COAL'S SHARE OF THE ELECTRICITY MARKET HAS INCREASED 20 PERCENT WHILE THAT OF NATURAL GAS HAS PLUMMETED 60 PERCENT.

PART OF THAT REASON LIES IN RELIABILITY. FOR ALL OF THE HOOPLA SURROUNDING NATURAL GAS, INCLUDING THE INDUSTRY'S OWN CLAIMS TO BEING SUPERIOR TO COAL, THE FACT REMAINS THAT PROVEN NATURAL GAS RESERVES AT CURRENT USAGE RATES NOW STAND AT A MERE EIGHT-YEAR SUPPLY. THAT COMPARES TO WELL OVER A CENTURY FOR COAL. AND IF NATURAL GAS WERE TO COMPLETELY REPLACE COAL CONSUMPTION IN THIS COUNTRY--AS I ASSURE YOU SOME ENVIRONMENTALISTS WOULD DESIRE--THAT SUPPLY DWINDLES TO JUST FOUR YEARS' WORTH.

UTILITIES TODAY MIGHT BE ABLE TO GET A FIVE-YEAR CONTRACT FOR NATURAL GAS, AND THEY MAY HAVE SOME PRICE PROTECTION IN THE SHORT TERM. BUT I WOULD HATE TO BE THE CHIEF NATURAL GAS BUYER FOR A UTILITY TRYING TO PREDICT WHERE PRICES OR AVAILABILITY FOR THAT GAS WILL BE WHEN THAT CONTRACT RUNS OUT.

PRICE, OF COURSE, QUICKLY FALLS VICTIM TO SHORT SUPPLY. AND, AS A RESULT, WE HAVE ALREADY SEEN NATURAL GAS PRICES ROCKET 80 PERCENT FROM LEVELS OF A YEAR AGO. THESE AREN'T THE SORT OF NUMBERS THAT OFFER CONFIDENCE TO UTILITY PLANNERS AND FUEL BUYERS.

THESE NUMBERS, TOO, SHOULD NOT INSPIRE CONFIDENCE AMONG THE NATION'S DECISION MAKERS. AND IT IS OUR JOB TO MAKE SURE THEY ARE MADE AWARE OF THESE FACTS. OUR GOAL IS NOT TO TEAR DOWN THE NATURAL GAS INDUSTRY, BUT TO POINT OUT THAT IN TERMS OF RELIABLE, ABUNDANT, INEXPENSIVE DOMESTIC ENERGY, COAL STILL STANDS ALONE.

I DON'T KNOW HOW OUR ENERGY WILL BE SUPPLIED A CENTURY FROM NOW. IT MAY WELL BE FROM WIND OR GEOTHERMAL ENERGY OR SOME SOURCE NOT YET EVEN CONSIDERED. BUT THE HARD FACT IS THAT ONE ENERGY SOURCE IS IN A POSITION TO BEAR THE MAJOR BURDENS OF ADVANCING OUR CIVILIZATION. THAT ENERGY SOURCE IS COAL.

OUR CHALLENGE, THEN, IS MORE THAN A TECHNICAL ONE. WE MUST IMPROVE COAL'S REALITY, IF YOU WILL, THROUGH ADVANCED TECHNOLOGY. BUT WE MUST ALSO BRIDGE THE YAWNING GAP BETWEEN REALITY AND PERCEPTION.

AS JOHN PAUL NOTED IN HIS PLENARY SESSION REGARDING THE CEED PROGRAM, WE MUST BETTER COMMUNICATE THE STRONG EFFORTS OF THE PEOPLE IN THIS ROOM AND ELSEWHERE IN ADVANCING THE POTENTIAL OF THIS MOST NECESSARY FUEL THROUGH ADVANCED TECHNOLOGY.

WE MUST TAKE TO TASK THE JUNK SCIENCE ADVOCATES WHO WOULD HAVE YOU BELIEVE THAT THE SKY IS FALLING AND THE EARTH IS WARMING. WE MUST CHALLENGE THE FLAWED CONCEPT THAT ENERGY USE IS A SIN TO BE TAXED. AND WE MUST CONTINUE TO SHOW THAT IT IS IN THE NATION'S INTEREST TO AGGRESSIVELY PURSUE PRACTICAL, COST-EFFICIENT CLEAN COAL TECHNOLOGIES. THE PRESIDENT AND HIS ADMINISTRATION HAVE CALLED FOR A STRONGER PUBLIC-PRIVATE PARTNERSHIP IN PURSUING TECHNOLOGIES TO IMPROVE INDUSTRIAL COMPETITIVENESS. AND WE SHOULD CONTINUE TO EXPLORE THESE OPPORTUNITIES.

WE MUST ALSO COMMUNICATE THE MAJOR ROLE TECHNOLOGY HAS PLAYED IN ADVANCING THE CAUSE OF COAL IN THE PAST. FOR WHEN COAL'S FUTURE HAS LOOKED MOST BLEAK, TECHNOLOGY HAS NEVER FAILED TO LEAD TO BREAKTHROUGHS IN SAFETY, IN PRODUCTIVITY, IN EFFICIENCY AND IN THE ENVIRONMENT.

TODAY'S TECHNOLOGY HAS ASSISTED US IN <u>PRODUCTION</u>, WHERE WE HAVE IMPROVED PRODUCTIVITY BY 78 PERCENT SINCE 1970 AND DECREASED THE NUMBER OF MINING FATALITIES BY THE SAME PERCENTAGE.

TODAY'S TECHNOLOGY HAS IMPROVED <u>EFFICIENCIES</u>, ENABLING US TO OBTAIN THE SAME AMOUNT OF ENERGY FROM ONE TON OF COAL AS WE GOT FROM EIGHT TONS OF COAL EARLIER IN THIS CENTURY.

AND TODAY'S TECHNOLOGY HAS ENABLED US TO IMPROVE

ENVIRONMENTAL PERFORMANCE. SINCE 1970, SULFUR DIOXIDE OUTPUT

HAS DECREASED BY 27 PERCENT DURING A TIME WHEN AMERICA'S

ELECTRIC UTILITY COAL BURN INCREASED BY 144 PERCENT.

THERE IS ONE MAJOR DIFFERENCE BETWEEN NOW AND 1970 REGARDING THE CLEAN AIR ACT AMENDMENTS. AT THAT TIME, WE CALLED FOR MUCH GREATER RESEARCH INTO CLEAN COAL TECHNOLOGIES BECAUSE THERE WAS A MARKED VACUUM IN THIS AREA. TODAY, AS WITNESSED BY THE ATTENDANCE HERE, THAT VACUUM IS GONE. CLEAN COAL HAS ARRIVED.

WITHOUT GROWTH IN COAL USE, AMERICA'S POWER PRODUCTION WOULD BE LIMITED TO 1970 LEVELS. AND SO, MOST LIKELY, WOULD THE ECONOMY.

EACH \$1 BILLION WORTH OF COAL PRODUCTION PRODUCES \$25 BILLION OF ELECTRICITY, \$10 BILLION IN TRANSPORTATION ACTIVITY, AND \$27 BILLION IN BUSINESS SERVICES.

THESE DYNAMICS OCCUR WITHIN AN ECONOMY THAT, IN ORDER TO GROW, WILL REQUIRE ADDITIONAL ELECTRICITY USE. LAST YEAR'S NATIONAL ENERGY POLICY ACT, FOR INSTANCE, REFLECTS A 30 TO 60 PERCENT INCREASE IN POWER DEMAND BY THE YEAR 2010. AND IT ACKNOWLEDGES THAT COAL WILL BE REQUIRED FOR AT LEAST HALF OF THE NEW BASELOAD IN THIS COUNTRY.

IN SHORT, THESE CLEAN COAL TECHNOLOGIES NOW IN DEVELOPMENT ARE CRITICAL TO AMERICA'S FUTURE IN A WORLD OF TOO LITTLE RELIABLE ENERGY. ON A GLOBAL SCALE, THEY WILL BE NECESSARY TO THE SMOOTH OPERATION OF MATURE ECONOMIES, AND CRUCIAL TO MEETING THE NEEDS OF THE WORLD'S DEVELOPING COUNTRIES.

AND SO, DESPITE DAILY CRITICISMS, THE PROSPECTS FOR COAL ARE STRONG. THAT'S MY PERCEPTION, AND I BELIEVE THAT WILL BE THE REALITY.

NOBODY EXPECTED MUCH FROM THE COAL INDUSTRY IN 1970. MANY WERE WRITING OBITUARIES. YET COAL IN THE SUCCEEDING 20 YEARS GREW AS IT NEVER HAD BEFORE.

TODAY, THE INDUSTRY CONTINUES TO BE WILLING TO PERFORM THE HEAVY LIFTING FOR A NATION'S ECONOMY. AND WE CONTINUE TO TAKE ALL STEPS NECESSARY TO REMAIN THE NATIONS LOWEST-COST, MOST ABUNDANT FUEL SOURCE.

WE ARE ALSO COMMITTED TO BRIDGING THAT GAP BETWEEN PERCEPTION AND REALITY. THE COAL INDUSTRY AND OTHERS HAVE BEGUN THIS LONG AND DIFFICULT PROCESS OF CHANGING PUBLIC OPINIONS.

NEW TECHNOLOGIES, TOO, WILL BE TRIED BOTH IN THE MARKETPLACE OF COMMERCE AND THAT OF PUBLIC PERCEPTION. I WOULD URGE EACH OF YOU, AS TECHNOLOGY DEVELOPERS, TO JOIN US WHERE YOU CAN IN EACH OF THESE AREAS.

WE ARE FAST APPROACHING THE POINT WHERE ASPIRATIONS AND CONCERNS WILL HAVE TO BE RECONCILED; WHERE TALK IS SET ASIDE AND DECISIONS BEGIN TO FLOW; WHERE TODAY'S PROMISE BEGINS TO CROSS OVER INTO TOMORROW'S REALITY.

IF A STRONG ECONOMY AND GOOD JOBS ARE A GOAL, THEN ELECTRIC POWER FROM COAL WILL BE NECESSARY. AND YOUR CONTRIBUTIONS WILL ADD TO AMERICA'S COMPETITIVENESS WHILE IMPROVING THE NATURAL ENVIRONMENT.

IF ADDITIONAL CONSERVATION OF AMERICA'S ENVIRONMENT IS THE GOAL, THEN YOUR TECHNOLOGIES ARE THE MEANS OF ACHIEVING IT WITHOUT CLEAR-CUTTING THE ECONOMY.

THIS IS WHAT PROGRESS IS ALL ABOUT...THE MARRIAGE OF RESPONSIBLE CONSERVATION AND OF SOUND ECONOMICS.

IF WE SUCCEED, WHAT KIND OF WORLD COULD WE HAVE 20 YEARS FROM NOW?

IF TECHNOLOGY AND COAL ARE ALLOWED TO DO WHAT WE KNOW THEY CAN DO -- TO REMOVE THE UNDESIRABLE ELEMENTS FROM THE FUEL WHILE RETAINING ITS GOODNESS -- IT CAN BE A WORLD WHERE:

- -- AMERICA WILL NEVER AGAIN HAVE TO GO TO WAR TO PROTECT THE WORLD'S DOMINANT OIL RESERVES;
- -- WHERE ELECTRIC CARS HUM ALONG OUR HIGHWAYS, FREE OF POLLUTANTS AND CARBON DIOXIDE EMISSIONS;
- -- WHERE ELECTRIC-UTILITY RATES CAN BE FORECAST YEARS AND DECADES IN ADVANCE;
- -- AND WHERE THE REALITY OF INEXPENSIVE, RELIABLE DOMESTIC ENERGY CONTINUES TO DRIVE THE STRONGEST ECONOMIC MACHINE ON EARTH.

THAT'S THE WORLD I SEE. AND THAT'S WHY, TO ME, THERE IS NO QUESTION BUT THAT COAL REMAINS AMERICA'S FUEL OF CHOICE ... AND FUEL OF NECESSITY.

THANK YOU VERY MUCH.

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Clean Coal Technology Markets Session

Panel Chair:

Herbert Wheary,

Chairman, Utility Advisory Committee,

Southern States Energy Board



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EVOLUTION OF DOMESTIC UTILITY MARKET STRUCTURE INTO THE 21ST CENTURY

George T. Preston
Electric Power Research Institute

My comments focus on the evolution of the United States domestic electric utility market structure and some of the implications of that evolution for clean coal technology markets. I'll briefly address:

- recent and potential future changes in the electric utility industry
- projected U. S. electricity demand into the next century
- current and advanced coal-based electric generating technologies and their competition
- the domestic market for CCT electricity generation.

THE CHANGING ELECTRIC UTILITY INDUSTRY

The U.S. electric utility industry consists of over 3000 private and public companies and agencies with an aggregate power generating capacity of over 700 GW. This is the largest concentration of electricity capability in the world - larger than the next 5 countries combined. Of the U.S.' total generating capacity, 41% is coal-based, and in 1993 54% of our electricity will be produced from these plants.

The business environment in which the industry operates is changing rapidly.

- The customer is more sophisticated and more demanding.
 - Customers want more influence on the business direction of their utility.
 - Customers expect more breadth of choice in the services offered.

An industry that is used to having 100% market share has nowhere to go but down, so this new muscle flexing by customers requires a nimble response (Hayes, 1991).

- The composition of the industry the number and character of its participants - is changing.
 - Utilities are evaluating and deciding among a spectrum of organizational structures, ranging from the traditional vertically integrated to completely horizontally linked or separate unbundled organizations. The United Kingdom adopted the latter model - swallowing the whole pill in a very short transition time except for nuclear generation.
 - New players non-utility generators (NUGs), including independent power producers as well as those affiliated with regulated utilities - have entered the generation side of the industry and have accounted for over 50% of new generating capacity additions since 1990. This market share of capacity additions is likely to persist well into the first decade of the 21st century.
 - Several significant mergers and acquisitions have occurred or have been tried in the past few years, with more to come as utilities seek synergies to cut their fixed costs and remain competitive. Examples include PacifiCorp - Pacific P&L and Utah P&L; Centerior - Cleveland Electric Illuminating and Toledo Edison; Midwest Resources - Iowa Power and Iowa Public Service; Western Resources - Kansas P&L and Kansas G&E.
 - The "regulatory compact" is cracked, if not broken, as Alfred Kahn, a farsighted regulator, observed in 1988: "The industry also has been opened in various ways to unregulated competition, but very partially, and in ways that have given rise to all sorts of distortions, inefficiencies, and inequities Whichever path the future takes, the companies have every right and obligation to demand elimination of the distortions inherent in partial deregulation . . . " (Kahn, 1988).

• The regulatory framework is changing.

- The National Energy Policy Act has created new electric generation opportunities.
- Increased transmission access will broaden the market potential for IPP and APP (affiliated power producer) generation.
- Environmental regulation is still evolving, with increasing emphasis on pollutant prevention and externality-based cost incentives.
- Under integrated resource planning (IRP), many utility companies will not be able competitively to build, or even own, new generating capacity.

- The financial rules and corporate objectives are changing.
 - Electricity is still the product, but increasingly it is viewed by customers and the more perceptive companies as an energy service, not a commodity.
 - Corporate earnings come from multiple sources.
 - Corporate growth no longer depends on sales growth.
 - Sustained low interest rates are putting pressure on common stock dividends. (Wang, 1993)
- It is simplistic to say that any of these changes is driven unilaterally by any other. They all influence each other, but the corporate attitude toward electricity generation as a business is changing, driven by all of the above.
 - Generation is moving outside the rate base as IPPs and APPs account for over 50% of new capacity additions. However, most of the added capacity has been for peaking and cycling duty. Little baseload capacity will be added in the 1990s - meaning that installed baseload generation will continue to dominate electricity revenues.
 - For many reasons influenced by the driving factors cited above, IPPs and APPs tend to be the early implementers of new advanced generating technologies, out of proportion to their relative presence in the industry.
 - Utility corporate decisions about plant upgrades and maintenance investments will be determined by an asset management decision philosophy that looks beyond the "obligation to serve" and considers a broader definition of corporate value.

Economic life vs physical life. One implication of asset management based decisions is that the classic 30-year book life - assumed for many fossil generating plants at their commissioning - is becoming irrelevant. Plants can be designed and operated to have physical lives well beyond 30 years - even an "undefined" physical life; but if competition, downward price pressure and tightening environmental requirements along with technology advancements make a physically healthy but obsolete plant economically inoperable, then designing and maintaining it to be physically capable of a long life was not a viable business strategy. This is why the issue of relicensing nuclear plants has lost some urgency in recent years; even with years remaining on their 40-year licenses, several nuclear plants have closed. (Wang, 1993)

U. S. ELECTRICITY NEEDS

Growth in electricity demand will likely continue, since electricity is the most versatile energy source at the point of use. The U. S. Energy Information Administration's Annual Energy Outlook 1993 projects that electricity energy and load demand will increase at a 1.3 - 1.9% annual rate from 1990 to 2010 depending on the strength of the U. S. economy, the proportion of electricity relative to total U. S. energy consumption, the impact of higher efficiency industrial technologies and energy savings from demand-side management. Energy demand growth at even the lowest rate of 1.3% annually will require adding about 150 GW of new capacity between 1990 and 2010. This compares to an installed base of about 730 GW. A 1.9% growth rate implies about 250 GW of new capacity.

EIA and others expect that 50% or more of the generating capacity added between now and 2000 will be natural gas fired, to serve intermediate and peak load requirements. As reserve margins decline and existing base load capacity becomes more fully utilized toward the end of the decade, coal-based generation additions will likely become more significant - according to EIA, 36-62% of all capacity additions during 2000-2010.

Compared to the EIA projections of need, the announced plans of utilities and other electricity generators are relatively consistent in terms of types of capacity to be added, although the amounts of capacity on the drawing boards are far less than the EIA projected demand.

- The Power Engineering survey of North American utilities identifies 69 GW of planned additions, of which 30 GW is coal, 15-18 GW gas, and 11 GW nuclear. The largest planned coal-fired units are 675-720 MW, and most of these show startup dates after 2005 (Smock, 1993).
- NERC data show planned U.S. (48 states) additions for 1993-2001 of 73 GW including 8.5 GW coal-fired, 40.7 GW oil or gas-fired by utilities (fossil steam, combustion turbine and combined cycle) and 14.2 GW by NUGs.
- Utility Data Institute shows 1990-2000 planned U. S. additions totaling 113 GW: 52.5 GW utility including 12.4 GW coal-fueled, and 60.6 GW non-utility including 11 GW coal-fueled (UDI, 1993).

Some of the data are net of annual planned plant retirements; but as implied earlier, a significant number of plants are likely to be retired early due to competitive pressures shortening their economic life. And these "early retirements" generally have not been reflected in utility forecasting (Wang, 1993).

U. S. UTILITY GENERATING TECHNOLOGIES, 1993 TO?

Investment decisions and, indirectly, the structure of the U. S. domestic utility market itself will be affected by the technological success of numerous development and demonstration programs now in progress.

Conventional fossil steam boilers. As implied earlier, the bulk of the electric generating capacity running in 2000 is running today, much of it baseloaded. Until recently the presumption has been that existing baseload capacity would be the benchmark for generating technology performance as well as economics. However, as explained earlier, new legislative and regulatory approaches (e.g. externalities, renewable energy production credits) and advanced lower-cost technologies could drastically shorten the economic life of much of this existing capacity base.

State-of-the-art power plant (SOAPP). Modern materials, component designs and emission control technologies are the basis of advanced steam condition (4500 psi, 1050°F double reheat) supercritical coal-fired plants with thermal efficiency in the 39-42% range. These plants could exploit some of the flue gas clean-up technologies demonstrated in the early rounds of the DOE Clean Coal Technology program.

Pulverized coal combined cycle air turbine/steam turbine plant with thermal efficiency over 47%. This is high-efficiency developing technology with potential for significant capital cost reductions.

Coal gasification combined cycle with 2500°F combustion turbines. The consortium-funded 100 MW Cool Water demonstration in the mid 1980s was the cleanest coal-based generating plant ever to operate up to that time. Three major suppliers now offer commercial IGCC plants using 2300° F ("F series") combustion turbines.

Advanced pressurized fluidized bed combustion applies the design, operating and materials lessons learned from several early utility-scale AFBC commercial plants and PFBC demonstration plants to achieve efficiencies in the 44-46% range while side stepping hot-gas filter material limitations through clever cycle design. This is developing technology that will be tested by Southern Company with DOE and EPRI funding support.

Combustion turbine combined cycle. As discussed earlier, through much of the 1990s combustion turbines - first "heavy frame" and then aeroderivative machines - and advanced cycles based on combustion turbine concepts are expected to account for most new generating capacity. The 2500°F combustion turbines for these plants will be available by about 2000 to provide thermal efficiency of 54% (LHV) in combined cycle service. DOE's Advanced Turbine

Systems program is aimed at developing machines to reach combined cycle thermal efficiencies of 60% or more.

With 50-60 GW of planned combustion turbine additions in the next decade, the prognosis for long-term availability of gas at assured prices is an important factor. This can be summarized as:

- There is plenty of gas in the ground.
- Gas producers and distributors are confident of their ability to deliver. Some will sign 10 or 15 year ("long-term") supply contracts with specific escalation terms.
- The producer industry recognizes an issue concerning their ability to provide gas in the potentially required quantities at \$4/MBtu or less. Success in this will depend in part on technology advances to keep production costs from rising.
- Utilities that are adding significant combustion turbine capacity (and also IPPs, if they bear the risks) are not taking anything for granted, and many are buying gas storage capacity.
- There is an effective cap of about \$4/MBtu on the price of gas, because at that level, integrated coal gasification combined cycle economics can beat out natural gas combined cycle in many utility generation situations.

Distributed generation means modular units in the 10 kW to 2 MW size range to meet localized electricity demand and replace "economy of scale" with "economy of production." Examples include solar photovoltaic cell arrays, internal combustion engines, small gas turbines, fuel cells, and batteries. Distributed generation will not replace the need for future large-scale central-station generation; however, the utility business-strategic benefits of distributed generation will have major impacts on siting philosophy, rate making and the competitive environment.

MARKET FOR DOMESTIC COAL-BASED GENERATION

The recent galloping changes in the U.S. electric utility industry, projections of electric power needs for the next ten years, and perspectives on the status of the generation technologies to be available, support the following observations about the prospects for broad implementation of clean coal technologies in the domestic market.

- NERC projections indicate that utility coal-based generating capacity will be only 5.5 GW greater in 2002 than now, in contrast to gas-fired and dual fuel fired which will be a total of 50 GW greater in 2002 than now.
- The requirement for integrated resource planning (IRP will be required in 45 states by 1995) will add to the list of options to be considered i.e. it will open the competitive door to demand-side management, inter-utility power purchases, and plant refurbishments. Inter-utility power purchases facilitated by increased transmission access will make it more difficult for smaller utilities to stay in the generation business i.e. to add new generating capacity of their own, whether coal-based or other fuel source.
- In today's utility business environment, regardless of thermal efficiency, reliability or environmental performance, a clean coal technology that can be competitive only if its capital costs are levelized over a 30 year period, will not succeed. The half-life of technology advancement today is so much shorter that we must re-think everything we thought we knew about power plant investment horizons.
- In the 1990s and even after 2000, NUGs and the technologies that are suitable
 for distributed generation will hold the advantage of less risk through
 smaller-size capacity increments, compared to clean coal technologies or
 other coal-based options that depend on economy of scale to "make the
 numbers."
- Several key competitive issues face new coal-based technologies in the nearterm power generation market. These include credible demonstration, costs competitive with natural gas options, and capability to meet continually tightening environmental regulations and externality challenges.
- The capital cost for most current or advanced coal-based technologies is in the range \$1300-1700/kW - which at today's gas prices can't compete with natural gas fired plants that cost \$500-700/kW. The coal technologies become competitive when natural gas reaches a sustained price of \$4-\$5 per MBtu or when one or more of the technically attractive clean coal options are developed sufficiently to be offered at reduced capital costs. Either or both of these could occur after 2000.

CONCLUSION

The U. S.' enormous low-cost coal resource base will continue to provide over half of the nation's electricity well after year 2000. For the balance of the decade, however, due to competitive pressures and the shortening half-life of technology advances, the low capital cost of natural gas generation options will make gas the predominant fuel for new capacity additions or repowering. This provides a

window in which to demonstrate advanced high-efficiency lower cost coal-based generating technologies.

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OPPORTUNITIES IN INTERNATIONAL DEPLOYMENT OF CCTs

Barry K. Worthington
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(The comments of Mr. Worthington were not available at the time of publication.)



INTEGRATED RESOURCE PLANNING: ITS IMPACT ON SUPPLY-SIDE OPTIONS

Steven A. Fluevog Project Engineer Georgia Power Company

(The comments of Mr. Fluevog were not available at the time of publication.)



Talking Points for Clean Coal Conference- IPP Perspective

P. Chrisman Iribe
U.S. Generating Company
September 8, 1993

I. Background on IPP Industry and U.S. Generating

Historical Growth (1978 - 1993)

- IPP industry has grown from 0-6% of U.S. electric capacity in 15 years
- 7-10% of IPP industry coal based, over 70% is natural gas based
- U.S. Generating has over 1200 MW of coal fired projects in construction or operation all permitted in last 3.5 years.
- II. Market Trend in Coal Combustion Technologies
 - A. Distinct Consumer (utility is IPP customer) preference for low cost competitively procured electricity is pushing the historical new technology "test-bed" (i.e., the rate based utility) off the stage.
 - B. Societal pressures for cleaner and "smaller" electricity facilities (smaller scale cogen sites in urban air sheds make clean projects easier to permit) further limits growth in solid fuel combustion.
 - C. Typical cost advantage of solid fuel consumption even with clean-up has been offset by efficiency advances in combustion turbine technology.
 - D. Gas costs now will have to more than double in real terms from current level to give coal even the appearance of competitiveness.

III. Clean Coal Technology Commercialization Issues

- 1. Can we with existing technology make clean coal projects that:
 - are almost as clean as gas plants SO_x , NO_x and particulate. (In reality comparison of new plant emissions should not be made between fuels but compared to existing fossil plants whether oil, gas and coal that are in reality 2-5 times dirtier).
 - can use waste water and zero discharge systems

2. Problem areas are:

- High CO₂ emission
- Solid waste concerns (ash)
- Air toxics could be a problem

Note: Today, natural gas fired turbine generation is nearly twice as efficient and even with 60 days of No. 2 oil firing generates between 1/6 and 1/4 the regulated pollutants as a coal fired facility (see table which follows).

IV. What needs to be improved if clean coal technology commercialization can go forward

- 1. Need to improve efficiency of use (e.g., gasification) and thus reduce CO₂ emissions
- 2. Need to develop safe, commercial opportunities to use ash
- Must continue to improve on particulate removal
- 4. Must do all of these without increasing capital costs

V. Potential Market today - next 5-10 years

- 1. Replacing older utility units (repowering) in domestic market
- 2. International in regions where there are limited gas infrastructure and/or substantial coal resources

VI. Commercialization Challenges - Conclusions

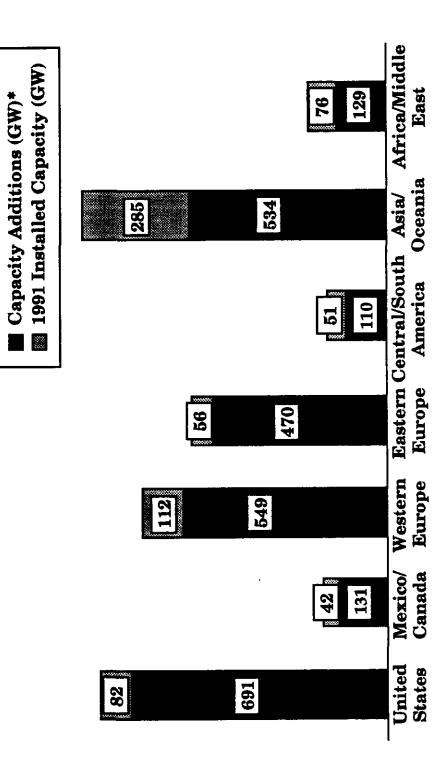
- 1. Loss of utility as test-bed for commercialization
- 2. IPP financing will inhibit commercialization of CCT
- Need to develop or find a mechanism for risk sharing with beneficiaries of the new technology i.e., major role for government, large trade associations, equipment and fuel suppliers
- 4. Project financing or lower equity commitments truly limits <u>all</u> but the surest technologies or the most profitable technology applications in order to offset commercialization risks

How To Get "Clean Coal" Into Foreign Markets?

- Make country or World Bank emission standards more stringent.
- Lower the cost of clean coal to be competitive with conventional technologies.
- Find "third party" sources of funding for the incremental cost of clean coal technologies.

World Power Markets

Installed Capacity and Planned Capacity Additions, 1991-2000 (Gigawatts)



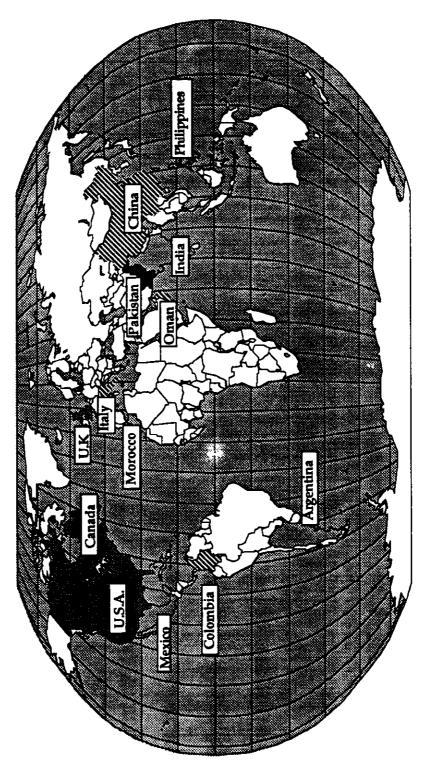
Note: Capacity additions include plant retirements and repowerings. Sources: RCG/Hagler, Bailly Inc., and American Tractabel

Environmental Performance of Coal Options

	Heat Rate (Btu/kWh)	NO _x (Ib/MMBtu)	SO, (Ib/MMBtu)	CO, (% of PC)
CONVENTIONAL PC	9550	0.5	4.4	100%
PC W/DRY SCRUBBER	9800	0.2	0.44	103%
PC W/WET SCRUBBER	10,100	0.2	0.22	106%
CFB	10,000	0.1	0.09	105%
IGCC	9200	0.1	0.04	%96

Assume: 2.5% sulfur; 11,500 Btu/lb coal.

Country Credit Risk

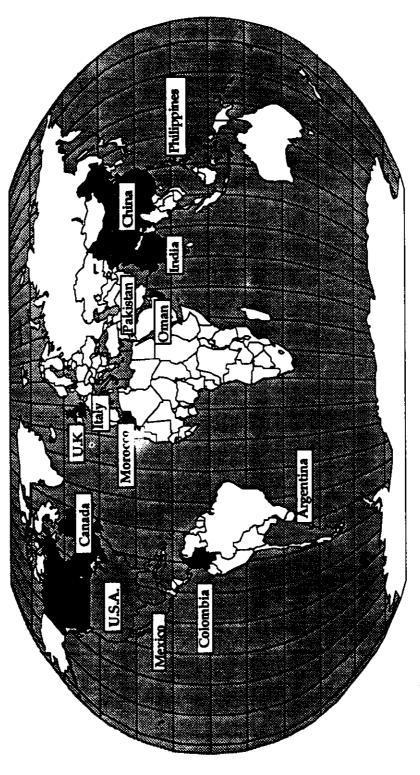


Few Risks Some Risks but Overall Quite Stable

Caution Risky

Source: Economist Intelligence Unit and Institutional Investors Credit Risk Ratings

(PP Potential Market Size*



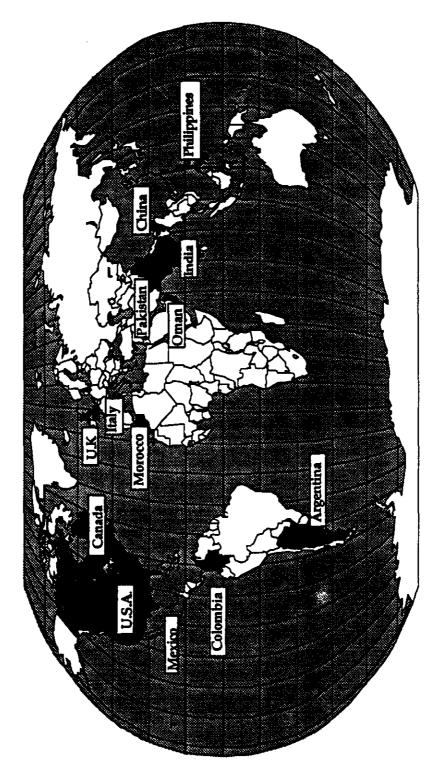
■ Large■ Medium

| Small

* IPP market size analysis includes planned capacity additions and planned asset sales

Source: International Private Power Quarterly

Jovernment Actions Encouraging IPP Development



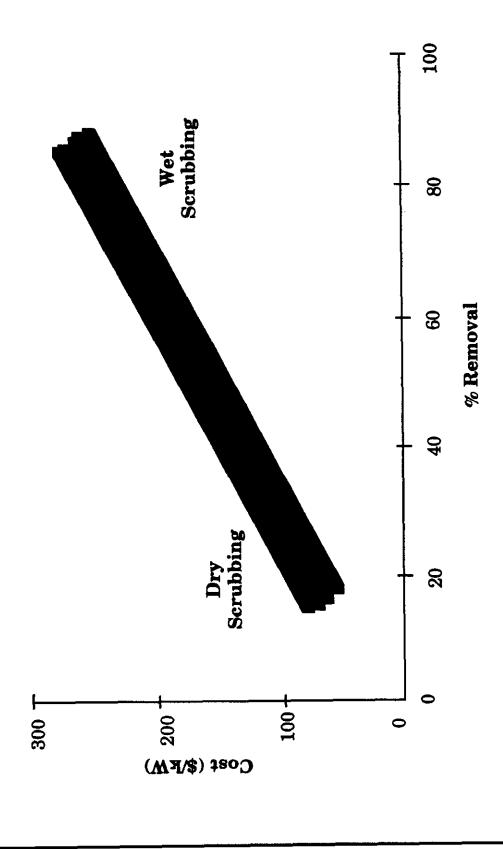
| Mature market, government encourages IPP development

Government encourages IPPs, but only to a limited extent

Immature market, few actions encouraging IPPs as of yet

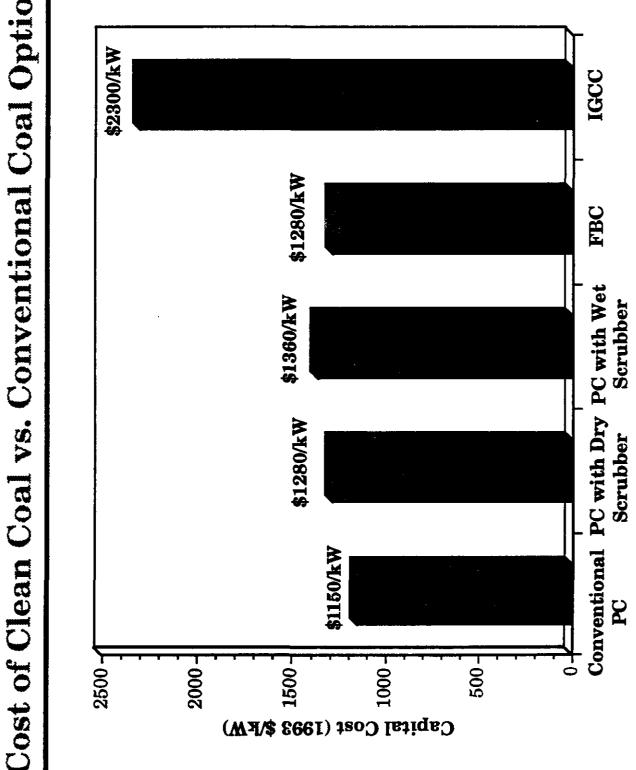
Sources: International Private Power Quarterly, AES project developers, and various other articles





Source: Battelle Labs. Includes both capital and NPV of operating costs.

Jost of Clean Coal vs. Conventional Coal Options



Source: AES experience with vendors.

A CASE STUDY: THE COMMERCIAL DEPLOYMENT OF PURE AIR'S CLEAN COAL TECHNOLOGY

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Second Annual
Clean Coal Technology Conference
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ABSTRACT

Northern Indiana Public Service Company (NIPSCO) has chosen an unique approach to comply with air quality regulations at its Bailly Generating Station. The utility has entered into a 20-year agreement with Pure Air to design, engineer, construct, fabricate, own, operate, maintain and finance the FGD project. Pure Air, a general partnership company between Air Products and Chemicals, Inc. and Mitsubishi Heavy Industries America, Inc., was selected by the U.S. Department of Energy (DOE) under the Clean Coal Technology Demonstration Program to install an advanced co-current, wet flue gas desulfurization (FGD) system at the Bailly Generating Station. The project combines the most advanced features of Mitsubishi's 95+ units worldwide (over 27,000 MW installed) and an innovative commercial arrangement into a single project to demonstrate substantially lower capital and operation costs when compared to conventional FGD designs. This paper briefly discusses the progress and performance of the project to date and then describes Pure Air's deployment strategy for this technology.

BACKGROUND

In the spring of 1988, with Clean Air legislation soon to be enacted, Northern Indiana Public Service Company (NIPSCO) and Pure Air, a general partnership of Air Products and Chemicals, Inc. and Mitsubishi Heavy Industries America, Inc., began discussions to determine what role flue gas desulfurization (FGD) could play in helping NIPSCO achieve compliance with the anticipated new SO₂ emission standards. The two companies submitted a proposal to the U.S. Department of Energy (DOE) and subsequently this project was selected for \$63 million of funding under Round Two of the agency's Clean Coal Technology Demonstration Program.

Innovative FGD Ownership

In October of 1989, Northern Indiana signed a flue gas processing agreement with Pure Air, whose scope includes the following: design, engineer, fabricate, construct, finance, own, operate and maintain an Advanced FGD facility adjacent to the Bailly generating station. Pure Air also assisted in the development of gypsum sales options and development of the eventual gypsum contract as part of its services to Northern Indiana.

Project Objectives And Accomplishments

The fundamental objectives of the project, as originally outlined by NIPSCO and Pure Air, were to achieve the required SO₂ emission reductions and minimize waste production at the least cost. The goal was to realize cost savings of roughly 50 percent compared to conventional FGD approaches by employing the following:

W2650aWCB 8/19/93

- Single 600 MW module which will reduce costs. Use of a single 100% capacity absorber module will demonstrate that spare modules are no longer necessary due to the high reliability of the module design.
- Co-current, single loop absorber with in-situ oxidation producing high quality gypsum while operating with a wide range of high sulfur coals. Oxidation will be accomplished by an innovative air rotary sparger system.
- The FGD supplier will own and operate the plant for 20 years or more and provide ongoing performance guarantees which will reduce operating risk and cost to utilities and their customers.
- Sale of commercial grade gypsum to a wallboard manufacturer.
- Direct injection of powdered limestone.
- High sulfur dioxide removal efficiency up to 95%.
- Wastewater Evaporation System (WES) which will reduce water disposal problems inherent with many U.S. power plants.
- Multiple boilers to a single absorber module which will significantly reduce costs at power plants with multiple boiler units.

Additionally, NIPSCO, Pure Air, and the DOE are in the process of employing an additional feature using Pure Air's proprietary technology for producing PowerChipTM gypsum. PowerChip gypsum is an agglomerated product using typical gypsum produced from an FGD facility and which can be substituted directly for natural rock gypsum in wallboard and cement manufacture. This eliminates any capital investment for the use of FGD gypsum by the end user. Unlike, the "pelletizing" process employed in Europe, PowerChip gypsum can be produced economically [approximately \$2.50/ton (including capital) versus \$8-10/ton for pelletizing].

MARKET FORCES

When considering the flexibility that utilities are given in complying with the SO₂ emission reduction requirements of the Clean Air Act Amendments of 1990, it is clear that traditional approaches to installing air pollution control systems must be modified to successfully compete in this new market. The "command and control" philosophy inherent in the New Source Performance Standards regulations dictated air that pollution control systems be built and operated regardless of the cost per ton of SO₂ removed. The Clean Air Act and the focus on least cost planning in an increasingly competitive power industry require a low cost, low risk, reliable compliance strategy for achieving environmental objectives.

Just as the actual FGD system awards in Phase I were significantly below most expectations, the demand in Phase 2 will be a function of how cost-effectively FGD technology can compete with other compliance options. Least cost will become the overwhelming driving force in making compliance decisions, just as it is today in making decisions as to how to generate new power in a very competitive marketplace.

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SO₂ Emission Allowances

Emission allowance trading provisions allow the transfer of emission rights from facility to facility and from utility to utility or independent power producers. Estimates show that emission allowance trading has the potential of reducing the costs of achieving SO₂ emission reduction requirements by 25 percent or more. The trading system permits utilities and independent power producers to buy, sell, and bank allowances, which the EPA has allocated to individual utility generating units. This new type of trading commodity is a license that grants the bearer the right to emit one ton of SO₂ per year. As a commodity, emission allowances will become a product themselves, a form of currency. Each ton of SO₂ emitted by a facility will have a value in the sense that if it were not emitted it could have been traded or sold to another facility.

It may be more cost-effective, for example, for one generating unit to overcomply and credit or sell its excess emission allowances to another facility which, in turn, may find it less costly to buy allowances than to install an expensive control system or switch fuels. Because of the newness of the emission allowance approach, it will be important for utility commissions to establish some form of review and certification procedure so that power generators can reflect the value of such allowances in their compliance plans. Several Midwestern states have in fact already passed legislation directing their commissions to review and approve such compliance plans.

Cost Analysis

In developing a least cost strategy utilizing FGD technology, it is critical to assess the potential impact of all cost elements. The use of advanced technology, the potential derates of 10-15% by fuel switching, by-product utilization, and most importantly, generating and crediting the value of emission allowances, are key strategies in compliance costs. For example, analysis of the cost of building and operating an FGD system at a hypothetical 500 MW generating unit located in the Midwest, burning 4.5 percent sulfur coal and using advanced FGD technology with an own and operate arrangement, by-product sales and emission allowances is shown in Exhibit 1. A graph depicting the impact of each element and a relative comparison to fuel switching is shown in Exhibit 2. The cost of building and operating a traditional FGD system would be over 50% higher than the compliance cost, which can be achieved by combining the potential savings of each element.

The cost per ton of SO₂ removed based on achieving 95% SO₂ removal efficiency, selling gypsum by-product at \$2/ton and selling or crediting emission allowances at \$300/ton, is calculated at \$236/ton SO₂ which is equivalent to a fuel delta of \$0.80/MMBtu. By comparison, the cost of using a conventional FGD system removing 90 percent, making a disposal grade by-product, and without crediting the value of allowances is \$373/ton SO₂ or \$1.26/MMBtu. The reduction of costs which can be achieved by combining the savings of each of these factors is not only important to optimizing the cost of using FGD technology, they are essential to determining whether or not FGD is the least cost compliance alternative. In order for our hypothetical

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Midwest generating unit to switch to compliance grade fuels, the plant would likely need to abandon the use of local high sulfur coal which it was designed to burn and import low sulfur coal from the West or Southern Appalachian coal regions. Coal price forecasts indicate that the cost delta for low sulfur compliance coal delivered to a Midwestern generating station will run approximately \$0.70/MMBtu on a 30 year levelized basis versus the cost of burning local coals in such units. The transportation delta alone accounts for 50 percent of this differential. In addition, even minor plant retrofits such as precipitator upgrades required to allow the burning of low-sulfur coals would increase the levelized cost to \$0.85/MMBtu. This analysis would indicate that without combining the benefits of advanced technology, by-product utilization and emission allowances, it is likely that fuel switching would be a lower cost compliance strategy.

Looking at the sensitivity of key cost variables such as the value of emission allowances, the sulfur content of the fuel burned, and the impact of landfilling gypsum by-product show a substantial change in the cost per ton of SO₂ removed, but demonstrate that combining the cost savings potential of each element is still essential to achieving the least cost compliance strategy. Exhibit 3 shows the cost per ton of SO₂ removed drops to approximately \$175 per ton if excess allowances were valued at \$600 per ton versus \$300 per ton. Exhibit 4 shows the cost per ton of SO₂ removed increases to approximately \$425 per ton if the sulfur content of the coal were 2% versus 4.5%. Exhibit 5 shows that landfilling by-product at a disposal cost of \$8 per ton increases the cost per ton of SO₂ removed to approximately \$275.

Least Cost Implications

The implications of these factors are equally important to retrofit and new plant markets, since the cost of achieving SO₂ emission requirements cannot be viewed simply in terms of the cost of installing and operating a mandated control technology. Use of low sulfur fuels, use of control technologies, and the purchase of emission allowances will all be viable, cost effective compliance alternatives. Along with the cost and performance risks of building, financing, operating, and maintaining an air pollution control system, the cost or value of buying, selling, or transferring emission allowances will become a critical factor in making FGD a least cost compliance alternative. The ability of suppliers to provide more than just equipment may become a key determinant in the ability of the marketplace to capitalize on the potential value of these factors.

It is likely that the provisions of the new Clean Air Act legislation will over the long term drive the marketplace for FGD systems to develop a least cost approach to SO₂ compliance which will incorporate many of the following factors:

- Reduced capital and operating costs through use of advanced technology.
- Third party financing, ownership, operation, and maintenance alternatives to capitalize on specialization, risk reduction, and economies of scale.

- Production and sale of commercial gypsum by-products.
- Creation and credit, lease, or sale of emission allowances from high removal (95 percent plus) systems.

Least cost for control technologies and all other compliance alternatives will be measured on a total cost basis expressed in terms of dollars per ton of SO₂. By capitalizing on the opportunities to reduce the capital and operating costs of FGD systems and generating excess emission allowances, the potential exists to meet or exceed the expectations of achieving the Clean Air Act Amendments requirements for SO₂ emission reductions at costs 25% lower than those which would have been incurred with a traditional "command and control" mandate. The ability of power producers, system suppliers, utility commissions, and fuel suppliers to work together to create and implement innovative strategies will be essential to capturing the full potential of the opportunities provided by this legislation.

DEPLOYMENT

The deployment of any Clean Coal Technology process has evolved beyond the standard competitive bid, turnkey methodology. The concept of "Allowances" embodied in the Clean Air Act Amendments of 1990 increases the flexibility and complexity of any SO₂ reduction decision making process by a U.S. electric utility. The strongest competitors to Pure Air are in reality non-scrubbing alternatives available to utilities. As discussed earlier, the concept of Least Cost in the absence of "Command and Control" regulation creates whole categories of decisions. Additionally, the value of Allowances and externalities, such as future costs of disposal, are ever increasingly being taken into account in a Least Cost analysis. Pure Air intends to deploy our technology to not only those utilities with SO₂ emission reduction requirements necessitated by Phase II of Acid Rain but also those utilities contemplating the addition of base loaded coal-fired generating capacity.

The former group is deciding between being a buyer or producer [for selling or banking] Allowances. Once they have decided to be a producer of Allowances they must determine whether to purchase low sulfur coal or SO₂ reduction technology. By packaging large, highly efficient AFGD systems, the taking of risk of gypsum sale and/or disposal and limestone procurement and with the Own-and-Operate concept Pure Air is offering a long-term least cost solution to a utility. Allowances and their future value will play a significant role in the actual decision and will remain an on-going parameter in the operation of any Acid Rain FGD system. As the value of an SO₂ Allowance ebbs and flows, a utility can choose to produce Allowances or emit SO₂ based on marginal costs.

The latter group of utilities will be competing with gas-fired IPP's for the construction [and inclusion in their capital rate base] of coal-fired, base loaded capacity. An FGD system will be required under "Command and Control" regulation and will represent a

major portion of the total cost of a grassroots coal-fired power plant. By employing the above mentioned concepts Pure Air can reduce the cost of a coal-fired, base load plant to assist the utility in making a least cost decision that allows them to construct their own coal-fired capacity. Without the employment of extreme measures most coal-fired, base loaded capacity that is required to compete with gas-fired IPP capacity will not be constructed due to current market prices of gas and gas-fired IPP capacity. Thus, once again the real competitor is a utility decision process not an alternate FGD vendor.

Lastly, a market is developing on the gulf and eastern coasts for fuel conversion of under-utilized oil-fired capacity to base loaded Orimulsion-fired operation. Orimulsion fired units will require SO₂ reduction and because these units do not have significant Allowances, highly efficient AFGD systems will be necessary. Due to the nature of these conversions from oil to Orimulsion, fuel savings will go to the benefit of the ratepayer while the risk of any capital expenditures will flow to the shareholders. Consequently, by incorporating the capital and operating costs (i.e., Own-and-Operate) into the cost of the fuel by either the fuel supplier or an other third party, the risk can be removed from the shareholders while the conversion can take place to the benefit of the ratepayer. This type of project can significantly reduce the average cost of production for a utility thus making them more competitive in their service territory. This will then bring benefit to their shareholders through increased power sales.

SUMMARY

As of this report, the facility is operating as expected. The AFGD facility has demonstrated sustained capability to remove in excess of 95% of the SO₂ from Units #7 and #8, has a 99.9% availability rate, and is producing a commercial-grade gypsum that is 98% pure, and being used to manufacture wallboard.

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TABLE 1
AFGD DEMONSTRATION TEST SCHEDULE

Test No.	Coal Sulfur	<u>Schedule</u>
1	2.0% to 2.5%	Fall 1994
2	2.5% to 3.0%	Fall 1993
3	3.0% to 3.5%	Fall 1992 (Complete)
4	3.5% to 4.0%	Spring 1993 (Complete)
5	4.0% to 4.5%	Spring 1994
6	Optimal Conditions	Spring 1995

TABLE 2
ADVANCED FGD PROJECT COST SUMMARY

	<u>Budget</u>	Actual/Estimate
Phase I (Design)	\$ 16,251,000	\$ 20,876,000
Phase II (Construction	\$ 93,142,000	\$ 85,654,000
Phase III (Operations	\$ <u>41,104,000</u>	\$ <u>43,067,000</u>
Subtotal	\$ 150,497,000	\$149,597,000
PowerChipTM Gypsum	\$ <u>1,210,898</u>	\$ <u>1,210,898</u>
Total	\$151,707,898	\$150,807,898

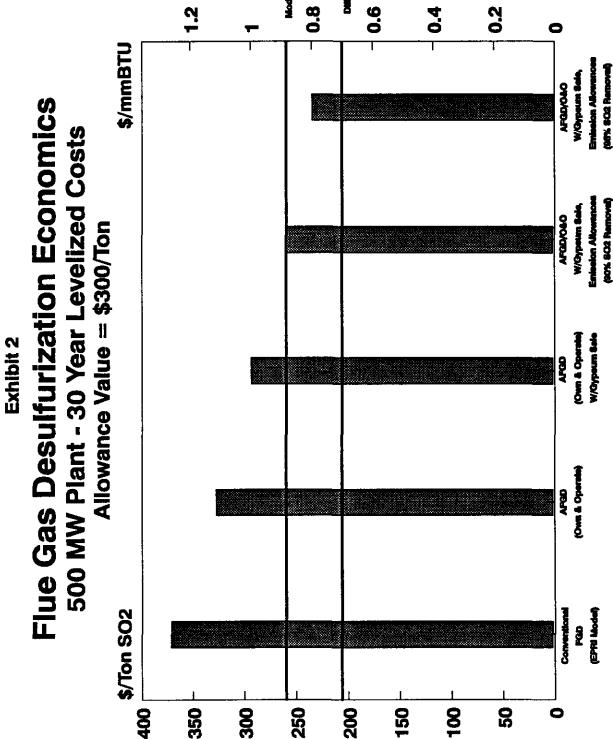
TABLE 3 OPERATIONS SUMMARY FOR PURE AIR SCRUBBER AT BAILLY STATION

	Expected	<u>Achieved</u>
SO ₂ Emissions	90% removal or 1.2 lb/MMBtu, whichever is less stringent	Averaged 95% (during DOE test up to 98+%, or 0.382 lb/MMBtu)
Power Consumption	•	ŕ
24-hour average	≤8,650 kW	5,962 kW
instantaneous	<9,650 kW	6,128 kW
Facility Pressure Drop		
24-hour average	≤13.5 IWC	6.66 IWC
instantaneous	≤14.5 IWC	7.55 IWC
Particulate Emissions	No net increase	0.04 inlet
(g/SCFD)		0.0071 outlet

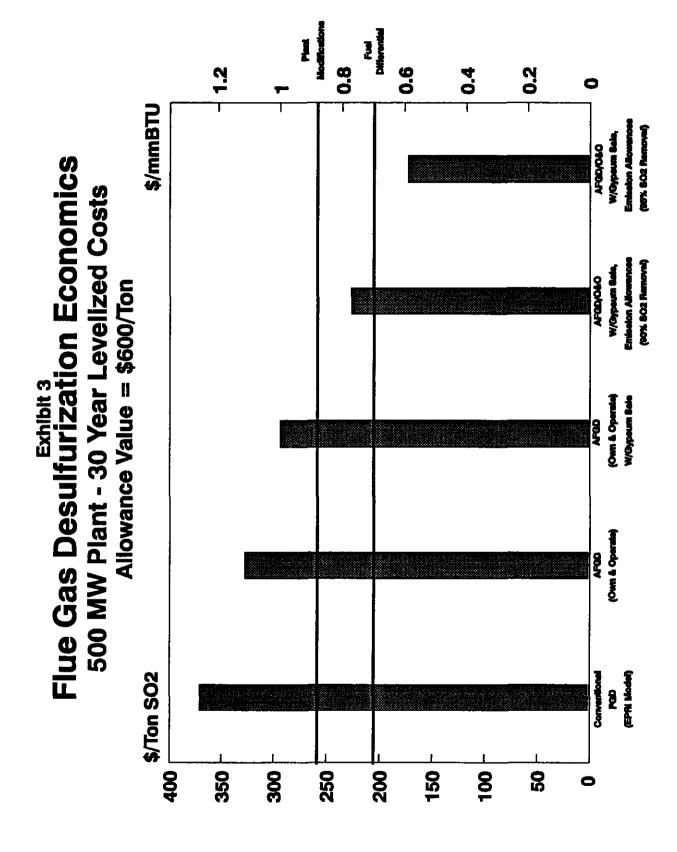
500MW FGD Economics (Capital + O&M) Allowance Value = \$300/Ton

Annual Costs (\$MM) \$/Ton SO2 \$/mmBTU 1.26 1.00 .88 8 30 Year Levelized Costs 295 236 261 373 329 29.8 37.6 33.3 26.4 23.9 **Emission Allowance Sale Emission Allowance Sale** (90% SO2 Removal) (95% SO2 Removal) Conventional FGD* (EPRI Cost Model) Own & Operate, Byproduct Sale, **Byproduct Sale,** Own & Operate, Own & Operate, Advanced FGD, Advanced FGD, Advanced FGD, Advanced FGD, **Byproduct Sale** Own & Operate (1990\$)

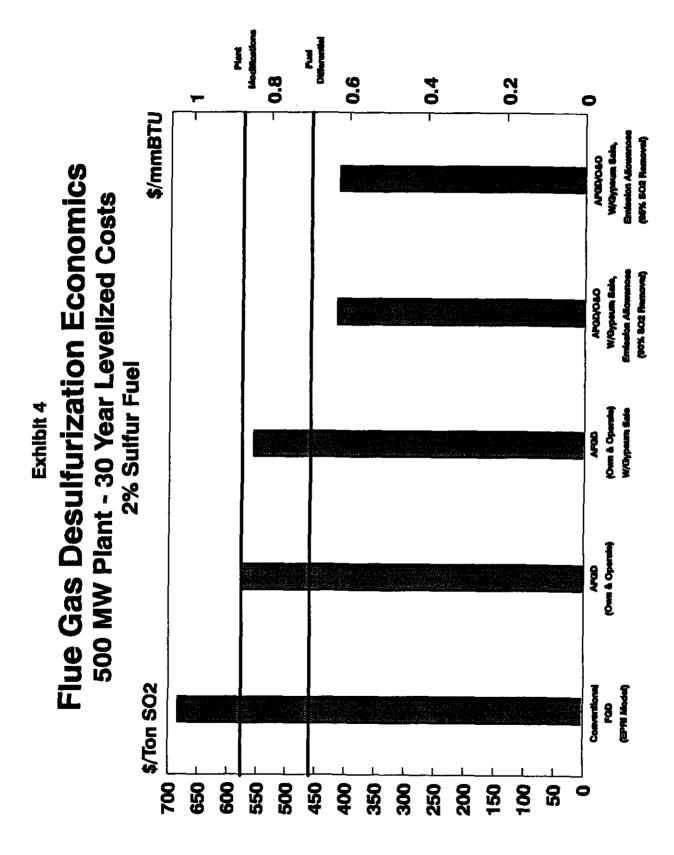
*Derived using EPRI's "Retrofft FGD Cost Estimating Guidelines", March 1990



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IMPACT OF FEDERAL ENERGY POLICY ON UTILITY PLANNING

Ray Billups
Manager, Industry Structure Issues, Governmental Affairs
Southern Company Services, Inc.

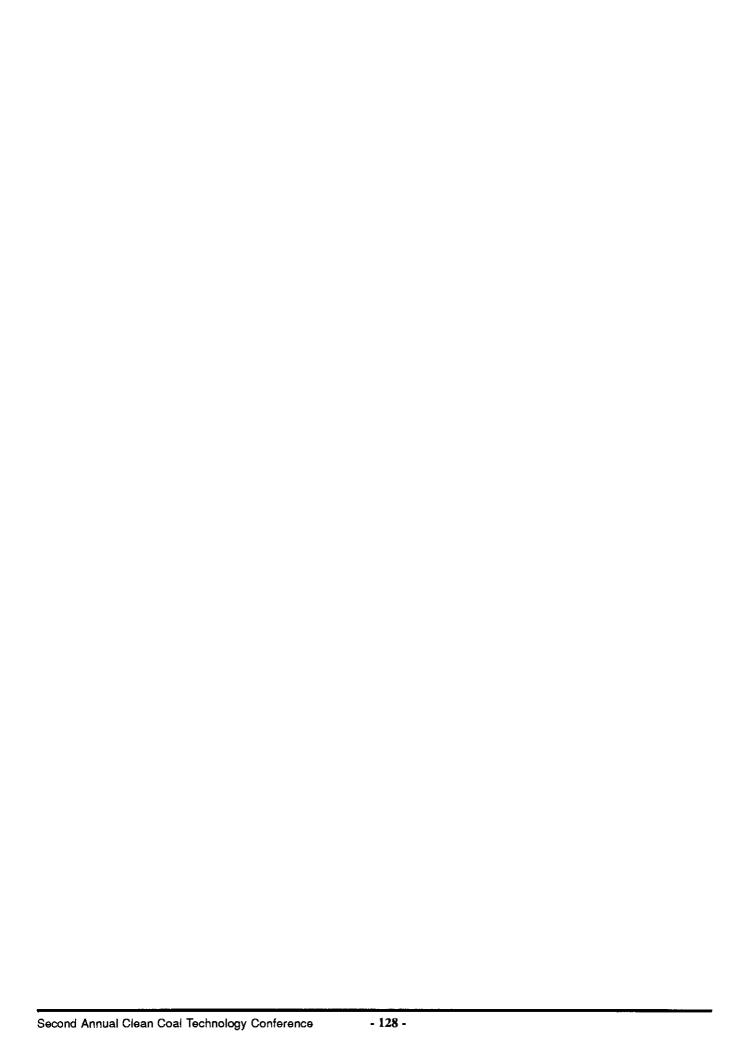
(The comments of Mr. Billups were not available at the time of publication.)



Session 1 NO_x Control Technologies

Co-Chairs:

Arthur L. Baldwin,
Pittsburgh Energy Technology Center/
U.S. Department of Energy
William E. Fernald,
Office of Clean Coal Technology/
U.S. Department of Energy



PERFORMANCE AND OPERATING RESULTS FROM THE DEMONSTRATION OF ADVANCED COMBUSTION TECHNIQUES FOR WALL-FIRED BOILERS

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ABSTRACT

This paper discusses the technical progress of a U. S. Department of Energy Innovative Clean Coal Technology project demonstrating advanced wall-fired combustion techniques for the reduction of nitrogen oxide (NOx) emissions from coal-fired boilers. The primary objective of the demonstration is to determine the long-term performance of advanced overfire air and low NOx burners applied in a stepwise fashion to a 500 MW boiler. A 50 percent NOx reduction target has been established for the project. The focus of this paper is to present the effects of excess oxygen level and burner settings on NOx emissions and unburned carbon levels and recent results from the phase of the project when low NOx burners were used in conjunction with advanced overfire air.

TABLE OF ABBREVIATIONS

AOFA Advanced Overfire Air

ASME American Society of Mechanical Engineers

C carbon

CF/SF Controlled Flow/Split Flame

Cl chlorine

CO carbon monoxide

DAS data acquisition system

DOE United States Department of Energy ECEM extractive continuous emissions monitor

EPA Environmental Protection Agency
EPRI Electric Power Research Institute

F Fahrenheit FC fixed carbon

FWEC Foster Wheeler Energy Corporation

H hydrogen

HHV higher heating value

ICCT Innovative Clean Coal Technology

lb(s) pound(s)

LNB low NOx burner LOI loss on ignition

(M)Btu (million) British thermal unit

MW megawatt
N nitrogen
NOx nitrogen oxides

NSPS New Source Performance Standards

O, O₂ oxygen

psig pounds per square inch gauge PTC Performance Test Codes RSD relative standard deviation

S sulfur

SCS Southern Company Services

SO₂ sulfur dioxide

UARG Utility Air Regulatory Group

VM volatile matter

INTRODUCTION

This paper discusses the technical progress of one of the U. S. Department of Energy's Innovative Clean Coal Technology (ICCT) projects demonstrating advanced combustion techniques for the reduction of nitrogen oxide (NOx) emissions from wall-fired boilers. This demonstration is being conducted on Georgia Power Company's Plant Hammond Unit 4, a 500 MW, pre-NSPS (New Source Performance Standards), wall-fired boiler. Plant Hammond is located near Rome, Georgia, northwest of Atlanta.

This project is being managed by Southern Company Services, Inc. (SCS) on behalf of the project co-funders: The Southern Company, the U. S. Department of Energy (DOE), and the Electric Power Research Institute (EPRI). In addition to SCS, Southern includes the five electric operating companies: Alabama Power, Georgia Power, Gulf Power, Mississippi Power, and Savannah Electric and Power. SCS provides engineering and research services to the Southern electric system. The ICCT program is a jointly funded effort between DOE and industry to move the most promising advanced coal-based technologies from the research and development (R&D) stage to the commercial marketplace. The goal of ICCT projects is the demonstration of commercially feasible, advanced coal-based technologies that have already reached the "proof-of-concept" stage. The ICCT projects are jointly funded endeavors between the government and the private sector in which the industrial participant contributes at least 50 percent of the total project cost. The DOE is participating through the Office of Clean Coal Technology at the Pittsburgh Energy Technology Center (PETC).

The primary objective of this demonstration is to determine the long-term effects of commercially available low NOx combustion technologies on NOx emissions and boiler performance. Short-term tests of each technology are also being performed to provide engineering information about emissions and performance trends [1]. Achieving 50 percent NOx reduction using combustion modifications is the goal of this project.

Following a brief unit and technology review, this paper focuses on (1) results of efforts to establish the relationship between NOx emissions and unburned carbon and (2) recent results from the low NOx burner (LNB) plus advanced overfire (AOFA) test phase.

UNIT AND TECHNOLOGY REVIEW

Georgia Power Company's Plant Hammond Unit 4 is a Foster Wheeler Energy Corporation (FWEC) opposed wall-fired boiler, rated at 500 MW gross, with design steam conditions of 2500

psig and 1000/1000°F superheat/reheat temperatures, respectively. The unit was placed into commercial operation on December 14, 1970. Prior to the LNB retrofit, six FWEC Planetary Roller and Table type mills provided pulverized eastern bituminous coal (12,900 Btu/lb, 33% VM, 53% FC, 72% C, 1.7% S, 1.4% N, 10% ash) to 24 pre-NSPS, Intervane burners. The burners are arranged in a matrix of 12 burners (4W x 3H) on opposing walls with each mill supplying coal to four burners per elevation.

During a spring 1991 unit outage, the Intervane burners were replaced with FWEC Controlled Flow/Split Flame (CF/SF) burners. In the CF/SF burner, secondary combustion air is divided between inner and outer flow cylinders (Figure 1). A sliding sleeve damper regulates the total secondary air flow entering the burner and is used to balance the burner air flow distribution. An adjustable outer register assembly divides the burner's secondary air into two concentric paths and also imparts some swirl to the air streams. The secondary air that traverses the inner path, flows across an adjustable inner register assembly that, by providing a variable pressure drop, apportions the flow between the inner and outer flow paths. The inner register also controls the degree of additional swirl imparted to the coal/air mixture in the near throat region. The outer air flow enters the furnace axially, providing the remaining air necessary to complete combustion. An axially movable inner sleeve tip provides a means for varying the primary air velocity while maintaining a constant primary flow. The split flame nozzle segregates the coal/air mixture into four concentrated streams, each of which forms an individual flame when entering the furnace.

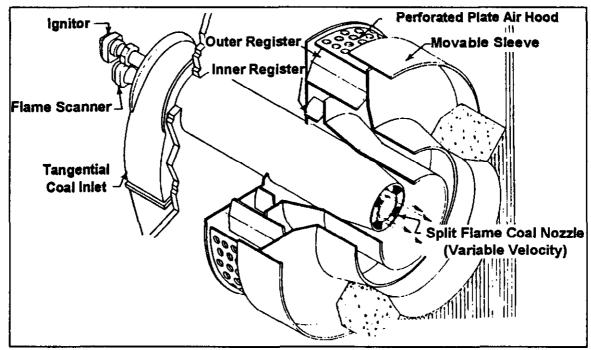


Figure 1. FWEC CF/SF Low NOx Burners

This segregation minimizes mixing between the coal and the primary air, assisting in the staged combustion process.

As part of this demonstration project, the unit was also retrofit with an Advanced Overfire Air (AOFA) system (Figure 2). The FWEC design diverts air from the secondary air ductwork and incorporates four flow control dampers at the corners of the overfire air windbox and four overfire air ports on both the front and rear furnace walls. Due to budgetary and physical constraints, FWEC designed an AOFA system more suitable to the project and unit than that originally proposed. Six air ports per wall were proposed instead of the as-installed configuration of four per wall.

During the course of the demonstration, the unit was also retrofitted with four Babcock & Wilcox MPS 75 mills (two each during the spring 1991 and spring 1992 outages). The unit is equipped with a coldside ESP and utilizes two regenerative secondary air preheaters and two regenerative primary air heaters. The unit was designed for pressurized furnace operation but was converted to balanced draft operation in 1977.

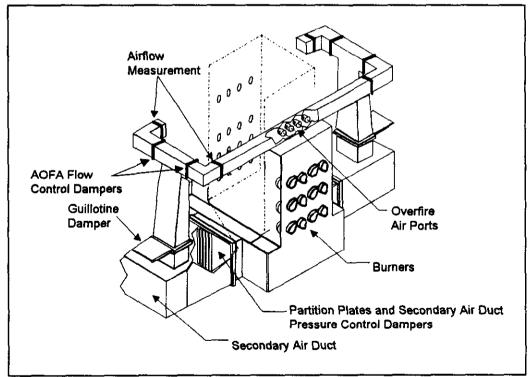


Figure 2. FWEC Advanced Overfire Air System

REVIEW OF PRIOR TESTING

Baseline, AOFA, and LNB test phases have been completed (Table 1). Short-term and long-term baseline testing was conducted in an "as-found" condition from November 1989 through March 1990. Following retrofit of the AOFA system during a four-week outage in spring 1990, the AOFA configuration was tested from August 1990 through March 1991. The FWEC CF/SF low NOx burners were then installed during a seven week outage starting on March 8, 1991 and continuing to May 5, 1991. Following optimization of the LNBs and ancillary combustion equipment by FWEC personnel, LNB testing was commenced during July 1991. However, due to significant post-LNB increases in precipitator fly ash loading and gas flow rate and also, increases in fly ash LOI which adversely impacted stack particulate emissions, the unit was run below 300 MW from September to November 1991 [2]. Following installation of an ammonia flue gas conditioning system, the unit was able to return to full load operation and complete the LNB test phase during January 1992.

Phase	Description	Date	
0	Pre-Award Negotiations		
1	Baseline Characterization	8/89 - 4/90	
2	Advanced Overfire Air Retrofit (AOFA) & Characterization	4/90 - 3/91	
3A	Low NOx Burner Retrofit (LNB) & Characterization	3/91 - 1/92	
3B	LNB+AOFA Characterization	1/92 - 8/93	
4	Digital Controls	9/93 - 6/95	
5	Final Reporting and Disposition	6/95 - 12/95	

Table 1. Project Schedule

Given the extended LNB test phase, insufficient time was available to complete the full requirements of the LNB+AOFA test phase prior to the spring 1992 outage; therefore it was decided to collect abbreviated data prior to this outage and comprehensive data following the outage. Following the outage, it was found that the AOFA had exacerbated the stack particulate emissions and the unit was again load limited, this time to 450 MW. While efforts were made to resume full load operation, special tests (i.e., NOx vs. LOI) were performed and long-term data collected. On March 30, 1993, Hammond Unit 4 resumed full load operation and comprehensive testing in the LNB+AOFA configuration began.

NOX VS. LOI TESTING

The NOx versus LOI testing was conducted between October 12 and 28, 1992. The primary purpose of these tests was to determine the effects of various burner settings and mill operation on NOx emissions and unburned carbon levels in the fly ash. To assess the effects of each parameter, the test matrix was designed so that a single parameter was varied each test day and all

other parameters were held constant to the extent possible. The parameters tested were (1) excess air, (2) mill coal flow bias, (3) burner sliding tip position, (4) burner outer register position, and (5) burner inner register position. The range of values tested is shown in Table 2. Mill characterization (i.e., primary air and coal through each mill; coal and air distributions; and particle size determination in each coal pipe) was also performed as part of this test program. Unless specified otherwise, all tests were run at the following conditions: (1) nominal 450 MW, (2) all mills in service with equal flows, and (3) overfire air flow set to 200,000 lb/hr (600,000 lb/hr of overfire air is normal for LNB+AOFA operation at this load). The tests were conducted at reduced loads to adhere to stack particulate compliance limits while overfire flow was maintained at the reduced level to prevent excessive slagging or overheating of the AOFA ports. Because of the different operating conditions (load and overfire air flow rates), the absolute values of emissions are difficult to correlate with previous test phase results; however, the intent of this test segment was to perform sensitivity studies, and the influence of the independent variables on NOx emissions and LOI at the tested condition should be indicative of the sensitivities at full load with LNBs and no overfire air.

		Range Tested		
Parameter Nominal Value		Low	High	
Excess Air	4%	2.8%	5.0%	
Sleeve Damper	7" Outer burner columns	Not	Not	
•	4" Inner burner columns	Adjusted	Adjusted	
Inner Register	~15%	Nominal	Nominal + 40%	
Outer Register	~60%	-20% of nominal	+20% of nominal	
Sliding Tip	+4 inches	+2 inches	+4 inches	
Mill Bias	No bias	Upper Mills +10%	Upper Mills -10%	
		Lower Mills -10%	Lower Mills +10%	

Table 2. Hammond 4 / NOx vs. LOI Tests / Parameters Tested

Figure 3 shows the range of the NOx and LOI values which resulted from this testing. NOx emissions and LOI levels varied from approximately 0.44 lb/MBtu to 0.57 lb/MBtu and 10 percent to 3 percent, respectively. With the exception of the excess O₂ tests, the NOx (in lb/MBtu) and LOI values shown in this figure are adjusted to a nominal 4 percent excess O₂ operating level using the slopes of the NOx and LOI vs. O₂ curves found during these tests. This adjustment was made to compensate for the test to test variations in excess O₂ levels. As expected, excess O₂ level had a considerable effect on both NOx and LOI (Figure 4). For the other parameters considered, within the range of adjustments tested, mill bias and sliding tip position had the greatest influence on NOx and LOI (Figures 5 and 6). As can be seen from these graphs, there is some flexibility in selecting the optimum operating point and making

tradeoffs between NOx emissions and fly ash LOI; however, much of the variation was the result of changes in excess O₂.

This can be seen more clearly in Figure 7 in which all the sensitivities are plotted. This figure shows for excess O₂, mill bias, inner register, and sliding tip, any adjustments to reduce NOx emissions are at the expense of increased LOI. In contrast, the slope of the outer register characteristic suggests that an improvement in both NOx emissions and LOI can be achieved by adjustment of this damper. However, due to the relatively small impact of the outer register adjustment on both NOx emissions and LOI, it is likely that the positive NOx / LOI slope is an artifact of process noise. It should be stressed that Figures 3 and 7 are parametric plots and that neither NOx or LOI are independent variables.

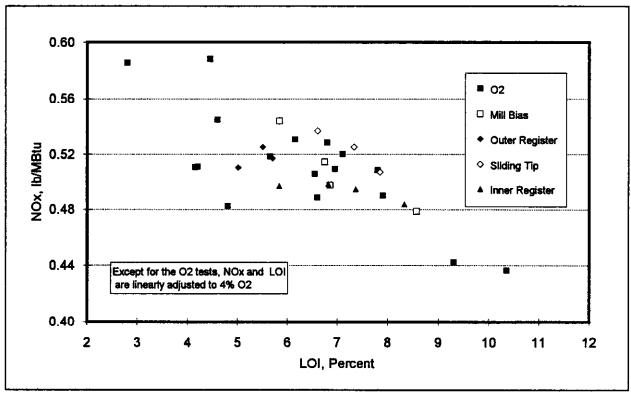


Figure 3. Hammond 4 / NOx vs. LOI Tests / All Tests

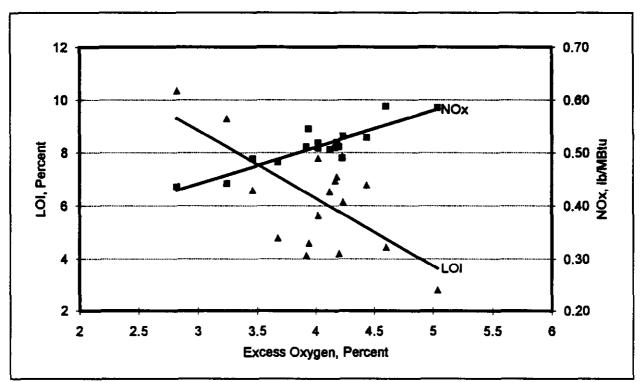


Figure 4. Hammond 4 / NOx vs. LOI Tests / O2 Sensitivity

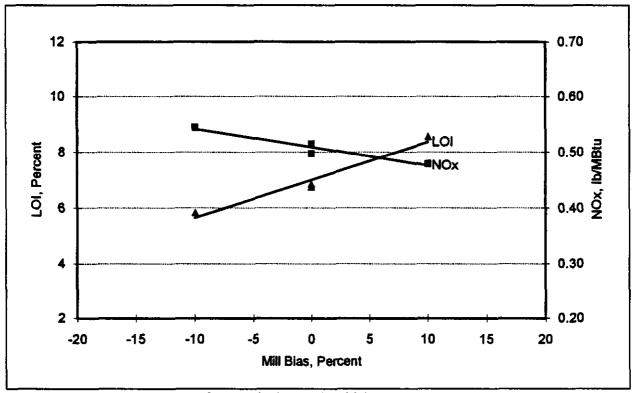


Figure 5. Hammond 4 / NOx vs. LOI Tests / Mill Bias Sensitivity

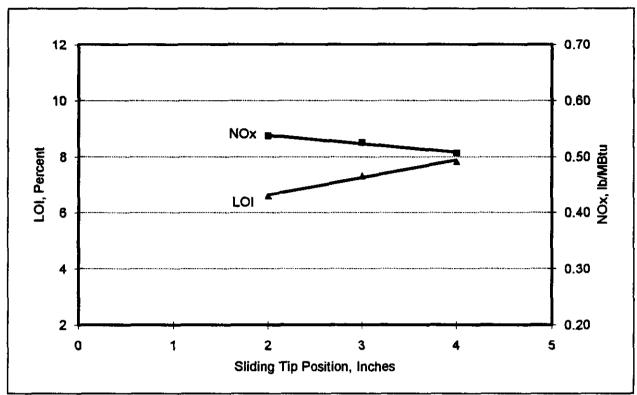


Figure 6. Hammond 4 / NOx vs. LOI Tests / Sliding Tip Position

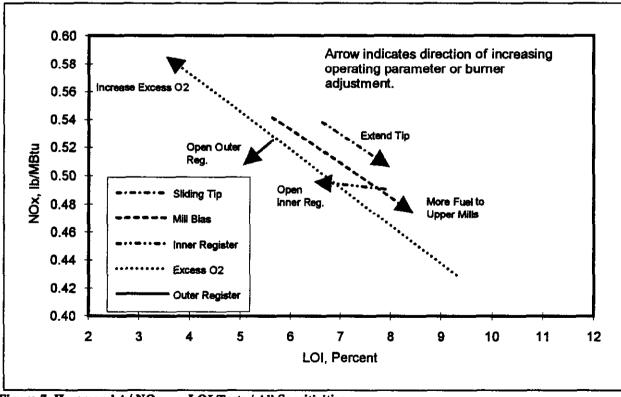


Figure 7. Hammond 4 / NOx vs. LOI Tests / All Sensitivities

LNB+AOFA CHARACTERIZATION

Following completion of the LNB test phase during January 1992, testing in the low NOx burner and advanced overfire air configuration was to begin with completion scheduled for late March 1992. However, due to delays associated with increased stack particulate emissions following the LNB installation, testing in the LNB+AOFA configuration could not be completed prior to the spring 1992 outage during which two new mills were to be installed. To obtain operating data prior to this outage, abbreviated testing (designated 3B') in the LNB+AOFA configuration was performed during February and March 1992. Following the spring 1992 outage, the unit ran at reduced loads (less than 450 MW) until spring 1993 to maintain stack particulate compliance. During this period, long-term data were collected and the NOx vs. LOI tests (discussed above) were performed.

Following resumption of full load operation on March 26, 1993, FWEC personnel re-optimized the unit starting March 30, 1993 and continuing through May 6, 1993. As shown in Figure 8, burner settings, with the exception of the burner tips, are similar to those used for the NOx vs. LOI test segment. The AOFA flow schedule is also shown in Figure 8. Since the AOFA is not automatically controlled, the operator must manually maintain not only the total overfire air flow rate but also balance the flows to the four corners of the AOFA windbox. This task has proven difficult during long-term, normal unit dispatch.

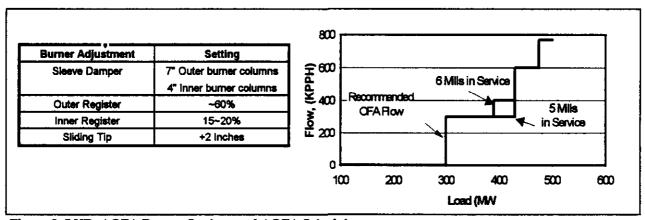


Figure 8. LNB+AOFA Burner Settings and AOFA Schedule

Subsequent to the re-optimization, comprehensive testing using LNB plus AOFA began. As of June 30, 1993, sixty-seven (67) diagnostic and performance tests have been conducted. As shown in Figure 9, full load NOx emissions are approximately 0.43 lb/MBtu with corresponding fly ash loss-on-ignition (LOI) values of 8 percent. At low loads (300 MW), NOx emissions and LOI are approximately 0.32 lb/MBtu and 5.5 percent, respectively. Also shown in Figure 9 are

the results from the February-March 1992 testing. NOx emissions for the latest round of testing are considerably below the NOx levels found in these earlier tests. The additional NOx reduction is most likely the result of re-optimization of the combustion system allowing lower excess air operation for the most recent testing (approximately 4 percent vs. 3.7 percent).

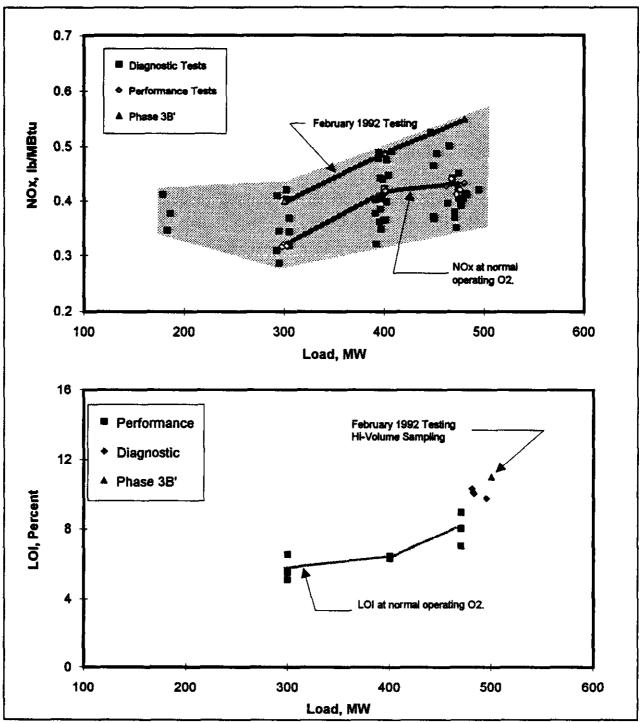


Figure 9. LNB+AOFA Short-Term NOx Emissions and Fly Ash Loss-On-Ignition

Long-term testing of the LNB+AOFA is in progress and is scheduled to continue until August 1993. As of June 30, 1993, twenty-nine (29) days of valid long-term data have been collected. Full load, long-term NOx emissions are approximately 0.42 lb/MBtu, which is consistent with that found during the performance testing (Figure 10). However, at 300 MW, long-term NOx emissions are near 0.37 lb/MBtu, nearly 0.05 lb/MBtu higher than the short-term emissions at the same load with approximately the same excess air and AOFA flow rate. The cause of this disparity is unknown. Despite this difference, the short-term data is within the 90th percentile range of the long-term data. As with the short-term data, a substantial difference exist between the current long-term NOx emissions and those previously recorded. This difference is again likely the result of re-optimization of the combustion system. Approximately 60 days of long-term data will be collected in this configuration; therefore, the final results may change when the complete data set is analyzed.

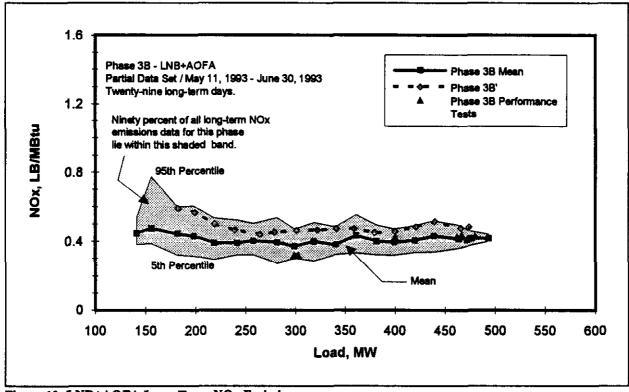


Figure 10. LNB+AOFA Long-Term NOx Emissions

DATA COMPARISON

As previously discussed, baseline, AOFA, and LNB test phases have been completed. Testing in the LNB+AOFA configuration is scheduled for completion in August 1993. The following paragraphs compare the results from these phases.

NOx Reductions

Figure 11 compares for Hammond Unit 4 not tuned for NOx aphases, following op to FWEC instruction long-term, full load, reduction averaged at the AOFA system de

Unit Configuration	Baseline		AOFA		LNB	
	Mean	RSD,%	Mcan	RSD,%	Mean	RSD,%
Number of Daily Averaged Values	52		86	-	94	•
Average Load (MW)	407	9.4	386	17.9	305	17.7
Average NOx Emissions (lb/MBtu)	1.12	9.5	0.92	8.6	0.53	13.7
Average O2 Level (percent at stack)	5.8	11.7	7.3	12.6	8.4	7.7
NOx 30 Day Achievable Emission Limit (lb/MBtu)	1.24	- 1	1.03	- 1	0.64	-
NOx Annual Achievable Emission Limit (lb MBtu)	1.13		0.93		0.55	-

Table 3. Long-Term NOx Emissions

LOI Performance

The fly ash loss-on-ignition (LOI) values increased significantly for the AOFA and LNB test phases and similar increases have been experienced in the LNB+AOFA testing (Figure 12). These LOI increases were evident over the load range. The LOI measurements were made during each performance test using EPA's Method 17 at the secondary air heater outlet [3]. As shown in Table 4, mill performance was generally better in the AOFA, LNB, and LNB+AOFA test phases than during baseline. The improvement in coal fineness was likely responsible for the reduction in fly ash LOI levels during the May-August 1993 LNB+AOFA test phase. Although it is commonly recognized that fuel fineness can have a pronounced effect on fly ash LOI, results from Plant Smith, Plant Gaston, and other sources indicate the direct impact of fuel fineness on NOx emissions is small [4,5,6]. As previously reported, the post LNB retrofit increase in fly ash LOI along with increases in combustion air requirements and fly ash loading to the precipitator, has had an adverse impact on the unit's stack particulate emissions [2].

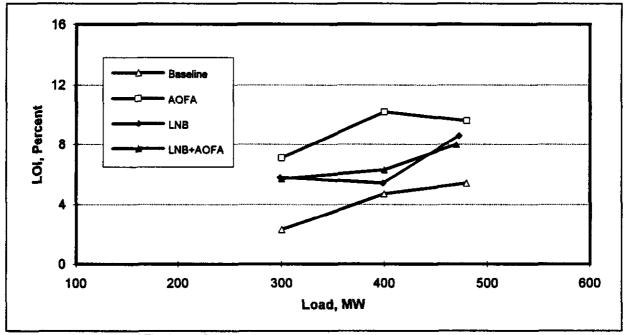


Figure 12. Hammond 4 / Fly Ash LOI

	Coal Fineness		
Technology	Passing 200 Mesh Percent	Remaining 50 Mesh Percent	
Baseline	63	2.8	
AOFA	67	2.6	
LNB	67	1,4	
LNB+AOFA	74	0.6	

Table 4. Hammond 4 / Mill Performance Summary

Excess O2 Levels

Long-term, economizer outlet O₂ levels for the AOFA, LNB, and LNB+AOFA test phases were generally higher than the corresponding baseline values (Figure 13). This change in O₂ level for these configurations is mostly attributable to an increase in combustion air requirements for the low NOx combustion configurations; however, factors unrelated to the retrofits, such as leakage in the furnace backpass, can also affect these levels. The impact of this leakage and varying O₂ levels on emissions and unit performance will be investigated and discussed in future reports.

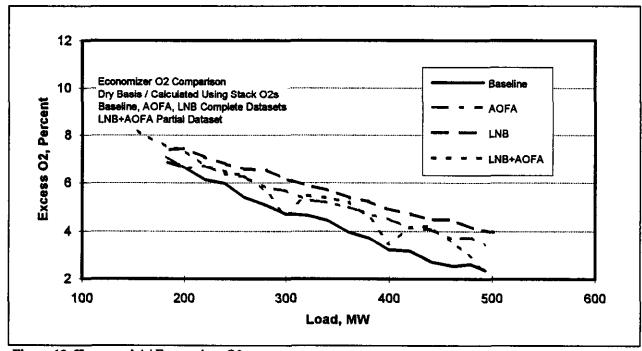


Figure 13. Hammond 4 / Economizer O2

CONCLUSIONS

In conclusion, the results to date at Plant Hammond indicate:

- NOx emissions have been reduced to about 50 percent of baseline values by using low NOx burners alone. These reductions were sustainable over the long-term test period and were consistent over the entire load range. At Hammond, preliminary results indicate AOFA used in conjunction with the LNBs provide approximately 15 percent additional NOx reduction benefit over LNB alone.
- For all low NOx combustion configurations, the unit experienced significant performance impacts including increases in excess air and fly ash LOI.
- At Hammond 4, operational and burner adjustments which favorably impacted NOx emissions adversely affected fly ash unburned carbon levels.

ACKNOWLEDGMENTS

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Measurement of Air Toxic Emissions from a Coal-Fired Boiler Equipped with a Tangentially-Fired Low NOx Combustion System

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ABSTRACT

This paper presents the results of measurements of chemical emissions from a coal-burning, tangentially-fired, utility boiler equipped with a hot-side electrostatic precipitator and a low NOx firing system. The tests were conducted in response to Title III of the 1990 Amendments to the Clean Air Act which lists 189 chemicals to be evaluated as "Air Toxics". The project was jointly funded by the Electric Power Research Institute and the U.S. Department of Energy under an existing Innovative Clean Coal Technology Cooperative Agreement managed by Southern Company Services. Field chemical emissions monitoring was conducted in two phases: a baseline "pre-low NOx burner" condition in September 1991 and in the LNCFS Level III low NOx firing condition in January 1992. In addition to stack emissions measurements of both organic and inorganic chemicals, plant material balance evaluations were performed to determine the efficiency of the hot-side ESP at controlling emissions of air toxics and to determine the fate of the target chemicals in various plant process streams.

ABBREVIATIONS

AAS atomic absorption spectroscopy

ABB CE Asea Brown Boveri Combustion Engineering Services

As arsenic

Btu British Thermal Units

C carbon or centigrade

Cl chlorine

Cr chromium

CVAAS cold-vapor atomic absorption spectroscopy

CVAFS cold-vapor atomic fluorescence spectroscopy

DNPH dinitrophenylhydrazine

DOE United States Department of Energy

EPA United States Environmental Protection Agency

EPRI Electric Power Research Institute

ESP electrostatic precipitator

F Fahrenheit or fluorine

FC fixed carbon

GC/MS gas chromatography / mass spectroscopy

H hydrogen

Hg mercury

HGAAS hydride generation atomic absorption spectroscopy

HHV higher heating value

ICCT Innovative Clean Coal Technology

ICP inductively coupled argon plasma emission spectroscopy

K potassium

lb(s) pound(s)

LNCFS Low NOx Concentric Firing System

ABBREVIATIONS (continued)

m meter

max maximum

min minimum or minutes

N Newton or nitrogen

NOx nitrogen oxides

O oxygen

P phosphorous

PAH polycyclic aromatic hydrocarbons

PETC Pittsburgh Energy Technology Center

PISCES Power Plant Integrated Systems: Chemical Emissions Studies

POM polycyclic organic matter

ppm parts per million

S sulfur

SCS Southern Company Services

SRI Southern Research Institute

T-fired tangentially-fired

UV ultraviolet

VM volatile matter

VOST volatile organic sampling train

μg micrograms

INTRODUCTION

This paper provides recent technical results on the release of chemical emissions from a U. S. Department of Energy (DOE) Innovative Clean Coal Technology (ICCT) project test site demonstrating advanced tangentially-fired (T-fired) combustion techniques for the reduction of NOx emissions from a coal-fired boiler. During the project, all three levels of the ABB Combustion Engineering Services (ABB CE) Low NOx Concentric Firing System ¹ (LNCFS) were evaluated. Chemical emissions tests were conducted before and after the installation of LNCFS Level III. Testing for the project was conducted at Gulf Power Company's Plant Lansing Smith Unit 2 near Panama City, Florida.

The ICCT project was managed by Southern Company Services, Inc., (SCS) on behalf of the project co-funders: the DOE, The Southern Company, and the Electric Power Research Institute (EPRI). The chemical emissions tests were funded by EPRI and DOE and conducted by Southern Research Institute (SRI). In addition to SCS, The Southern Company includes five electric operating companies: Alabama Power, Georgia Power, Gulf Power, Mississippi Power, and Savannah Electric and Power. SCS provides engineering, procurement, and research services to The Southern Company. The DOE is participating through the Office of Clean Coal Technology at the Pittsburgh Energy Technology Center (PETC).

The primary objective of this demonstration project was to determine the long-term effects of commercially available low NOx combustion technologies for T-fired boilers. However, this paper focuses on the results of the measurement of chemical emissions. The emissions of primary concern are those being addressed by the EPRI PISCES (Power Plant Integrated Systems: Chemical Emissions Studies) program. Most of these species are found among the "Air Toxics" listed in Title III of the 1990 Amendments to the Clean Air Act. The PISCES air toxics list is shown in Table 1. The substances in the measurement inventory include metallic and nonmetallic elements and organic compounds. Sampling and analytical methods, the test results, and inconsistencies in the results are presented in this paper.

UNIT DESCRIPTION

Plant Lansing Smith Unit 2, owned and operated by Gulf Power Company, uses a T-fired boiler (aspect ratio = 1.5 width/depth) rated at 180 MW with the capability to provide loads of up to 200 MW. The boiler is a Combustion Engineering radiant reheat, natural circulation steam generator which came on line in 1967. It is designed for continuous indoor service to deliver 1,306,000 pounds of steam per hour at normal rated load, a pressure of 1800 psig, and a temperature of

1000°F at the superheater and the reheater outlets. Five CE-Raymond bowl mills equipped with exhausters at the outlet of each mill deliver pulverized coal (66.5% C, 9.9% H₂O, 4.6% H, 1.4% N, 2.8% S, 6.3% O, 8.5% ash, 0.1% Cl; HHV = 11,886 Btu/lb, FC = 46.0%, VM = 35.6%) through 20 tangential coal nozzles with 5 nozzles stacked vertically in each corner of the furnace. The unit is equipped with Ljungstrom air preheaters and two forced-draft fans which deliver all the combustion air to the boiler. Exhaust gases are treated with both hot- and cold-side electrostatic precipitators (ESP). Although originally designed for pressurized furnace operation, the unit was converted to balanced-draft operation in 1976.

TEST DESCRIPTION

Chemical emissions were measured at Plant Smith on two occasions. Each test period required one week to complete. During the first period, tests were conducted with the LNCFS Level II technology in service. However, to simulate a baseline firing condition, the separated overfire air system was closed and the offset air nozzles were placed in line with the fuel nozzles. These baseline tests were conducted in September 1991.

During the second test period (January 1992), chemical emissions were measured with the LNCFS Level III in service. The LNCFS Level III technology is equipped with separated overfire air, close coupled overfire air, and offset air nozzles (Figure 1). During other portions of the test program, the long-term NOx reduction capabilities of the LNCFS Level III system were measured. At full load (180 MW), NOx reduction was 45 percent compared to the baseline emissions level (Figure 2). As unit load decreased, NOx emissions increased to baseline levels.

In each week of testing, samples were collected during two separate modes of ESP operation. For each test period, three tests were conducted with only the hot-side ESP energized and one test was conducted with both the hot- and cold-side ESPs energized. Each test required from 10 to 16 hours to complete.

The goals of the chemical emissions tests were to obtain the information required to answer the following questions:

- How are chemical emissions altered by the LNCFS Level III?
- How effectively does the hot-side ESP control chemical emissions?
- How much additional reduction in chemical emissions takes place when the cold-side ESP is energized?

The sampling plan was designed to include material balance checks of elements in fuel and discharge streams throughout the plant as well as in input and output streams across the ESPs and air heater. Discharge streams include the pyrite rejects, bottom ash, part of the bottom ash sluice water, economizer ash, ESP hopper ash, and stack gases. The sampling locations are diagrammed in Figure 3.

AIR TOXICS SAMPLING AND ANALYTICAL METHODS

Solid materials in bulk deposits (such as the ash discharged from the water-sealed furnace or ash deposited in hoppers) were collected at various intervals each day as grab samples. Daily composites of each material served for analysis. The individual solids collected for analysis included coal prior to pyrite removal, pyrite waste, bottom ash, economizer ash, and ESP ash. The coal composite was prepared from hourly samples from each feeder. The pyrite hoppers were inspected once per shift. All other solid samples were collected and composited once per day.

Gas streams entering the hot-side ESP or leaving the cold-side ESP (and then entering the stack) were sampled by methods developed by EPA or based on EPA sampling principles², and previously adopted as protocols for the PISCES program³. Table 2 lists the major sampling methods employed. This table also lists the collection media for the samples to be analyzed. An exception to EPA-based methodology was evaluated as an alternative method for sampling mercury in the vapor state. This method employed solid sorbents consisting of a quartz wool filter, followed by two KCl-soda lime traps, followed by two iodated carbon traps as recently described by Bloom⁴.

The analytical laboratories employed, in general, the methods that have been used in prior PISCES projects. Table 3 lists the analytical methods. Mercury from the solid sorbents was determined by cold-vapor atomic fluorescence spectroscopy (CVAFS), whereas mercury from the EPA train was determined by cold-vapor atomic absorption spectroscopy (CVAAS).

DATA ON METALS

Partitioning

Whether a trace metal occurs as a component of the ash or as a component of the gas phase is obviously an important factor insofar as control of its emission in an ESP is concerned. Significant conclusions with respect to this matter were possible, even though the Multi-Metals sampling train is limited in its ability to discriminate between the fractions of an element in the solid and gas phases. This limitation exists because the filter in the sampling train is maintained at

250 °F and thus it can easily collect an element that occurs as a vapor at a higher temperature in the duct being sampled. (The inlet gas temperature to the hot-side ESP was about 700 °F, and the outlet gas temperature from the cold-side ESP was about 325 °F.)

Two metals, mercury and selenium, were shown to be present predominantly in the vapor state at the outlet of the ESP, even given the limitation of the Multi-Metals sampling train. Mercury is volatile in various chemical states, both elemental and oxidized. The more conclusive evidence on the vapor state of mercury came from the samples collected with the solid sorbents, which explicitly avoided the collection of particulates. The fact that the concentrations of total mercury were comparable using the two methods provides complementary evidence of the vapor state. The concentrations of mercury found in the gas stream at the outlet of the ESPs ranged from 80 to 120 percent of the concentrations expected based on the mercury concentrations in the coal and firing rates of the coal. Selenium created persistent analytical problems, causing the material balance for selenium to be indeterminate. However, much of the selenium was found in the impingers behind the filter of the Multi-Metals train which substantiated a high volatility.

Arsenic is a metal that is appreciably volatile as the trioxide, and, in theory, might have been emitted from the stack in a vapor phase. In this study however, arsenic was shown conclusively to have been predominantly in the solid phase which was controlled by the ESP.

Concentrations of certain trace metals in ash samples that were separated from the gas phase at different temperatures indicated that metals other than mercury and selenium were in the vapor state before the gas reached the ESPs. Arsenic and antimony, for example, were much more concentrated on particulate filter samples taken out of the system at lower temperatures; presumably, therefore, they were in the vapor state at the higher temperatures.

Speciation

The chemical speciation, or oxidation state, of certain metals is of particular interest. In the case of mercury, emissions data regarding the ratio of the elemental form to the ionic form can be applied to plume chemistry and atmospheric deposition rates to provide insight on affected geographic locales. In the cases of chromium and arsenic, one oxidation state is considered to be very toxic, while a second is non-toxic or much more benign. However, in all cases of speciation measurements, the sampling and analytical procedures are still at various levels of development, and the potential for sampling artifacts is great. Many of the species display a wide range of measured concentrations, and probably a wide range of accuracy. All arsenic speciation data, for example, are especially suspect.

Mercury. As stated above, mercury is volatile in various chemical states. At 300 °F, for example, elemental mercury, Hg(0), has the highest volatility, while the chloride, HgCl₂, has a volatility that is just slightly lower⁵. Organomercury compounds, such as methylmercury, also have appreciable volatilities. Table 4 shows the distribution of mercury that was found in one of the sets of samples from the ESP outlet. The oxidized mercury, presumably HgCl₂, represented about 80 percent of the total, elemental mercury about 20 percent, and methylmercury only about 0.02 percent. The total concentration, 9.22 μg/Nm³, represents a material balance of 110 percent of the mercury supplied in the coal.

Chromium. Chromium in the hexavalent state is a carcinogen, while trivalent chromium is generally regarded as a non-toxic. The fraction of total chromium in ash samples that could be extracted in an aqueous alkaline medium and identified as Cr(VI) was determined by use of diphenylhydrazide as a calorimetric reagent⁶. In the ash entrained at the ESP inlet, 5-10 percent of the total chromium was in the hexavalent state. In the fine particulates that were not collected by the ESP but that remained entrained at the ESP outlet, the percentage of chromium measured in the hexavalent state was less definitely determined, but it appeared to be enriched in excess of 25 percent. However, the absolute concentration of Cr(VI) in the outlet stream from the ESP was very low since the removal efficiency for total chromium by the ESP was greater than 97 percent.

Arsenic. Arsenic can be toxic in both the trivalent and pentavalent forms. To the degree that the element could be extracted from ash in water, the quantities in the two oxidation states were determined by performing hydride generation atomic absorption spectroscopy (HGAAS) determinations on extracts acidified with citric acid (giving trivalent As alone) or with HCl (giving both oxidation states)⁷. Pentavalent arsenic was dominant in all the samples analyzed.

Control by Electrostatic Precipitation

With the cold-side ESP de-energized, the hot-side ESP in operation alone removed all but about 0.6 percent of the entrained fly ash during baseline testing. The hot-side and cold-side ESPs in combination gave no measurable improvement during the baseline testing. However, during the low NOx testing, the hot-side unit alone allowed a penetration of 1.0 percent compared to 0.6 percent for the combination.

Despite the predominance of most of the trace metals in the particulate phase, the observed ESP penetration by most of them was significantly more than 0.6-1.0 percent. Some examples of penetration on a percentage basis for metals that were predominantly in the particulate phase are as follows: arsenic, 1 percent; cobalt, 2 percent; manganese, 1-3 percent; molybdenum, 4 percent.

The finer ash particles appear to be enriched in the trace metals, probably as a result of metal deposition of surfaces at some time during transit of the gas stream from the furnace to the ESPs. Some removal of elements found in the gas phase at the ESP inlet at 700 °F might have occurred on cold surfaces in the air heater between the ESPs. However, this occurrence is not confirmed by the experimental data.

Material Balance

Overall system. The absence of information on mass flow rates of certain process streams, such as the rate of discharge of bottom ash, prevented a strict assessment of material balance. There was reason to believe, however, that for most of the metals the total elemental flow rate in the gas stream at the inlet to the hot-side ESP should have represented 80-100 percent of the elemental flow rate in the coal. The mass of ash entrained in this gas stream was approximately 80 percent of that in the coal (a figure in conformity with the rule of thumb that a pulverized-coal boiler will divide fly ash and bottom ash in an 80/20 ratio). There were no target elements that were profoundly enriched in the bottom ash. Thus, elements confined to the particulate phase should have been found at a level that was 80 percent of that supplied by the coal, and elements divided between the particulate phase and the gas phase, or found exclusively in the gas phase, should have been found at levels ranging from 80 to 100 percent.

Table 5 lists the ranges of trace metal "recoveries," that is, total elemental flow rates at the ESP inlet expressed as percentages of elemental flow rates in the coal. For the majority of the elements, the recoveries straddle the target value of 80 percent. In some cases, however, the range is so far biased from the expected range that the data cannot be truly said to represent recoveries. For example, such ranges as 131-256 percent and 26-46 percent for lead in the two test series reveal such serious analytical difficulties for coal and/or ash that neither set of results for lead can be regarded as meaningful. The inconsistencies almost certainly occur in the analytical procedures and not in recovery of a representative sample of the material entrained in the flue gas.

ESP system. Material balance could be determined more exactly insofar as the ESP system was concerned. Inlet and outlet mass flows in the gas stream were directly measured. The ash concentrations, corrected to 4 percent O2, were 7.29 g/Nm³ at the ESP inlet versus 0.038 g/Nm³ at the ESP outlet during baseline testing, and 7.73 g/Nm³ at the ESP inlet versus 0.079 g/Nm³ at the ESP outlet during the low NOx testing. The mass flow rate of collected hopper ash was not measured; but, it could be calculated as the difference between mass flows in the inlet and outlet ducts. Table 6 compares the trace element closures between the baseline and low NOx testing based on a ratio between the calculated accumulation rates of elements in the hoppers with the

difference between inlet and outlet duct flow rates. Generally, the closures across the ESP system signify better data quality than the recoveries at the ESP inlet.

Influence of Plant Variables on Emissions

The emission of trace metals was not affected in major ways either by converting the boiler to low NOx combustion or by operating with one or both ESPs. There was some evidence that low NOx combustion suppressed the fraction of total chromium that was present in the hexavalent state, which would be desirable.

DATA ON NON-METALS

Among the four non-metals considered, only phosphorus was found predominantly in the ash. In the fly ash deposited in the ESP hoppers, phosphorus in the form of P₂O₅ represented 0.22 percent of the total mass. In the coal ash obtained by laboratory ignition of the coal, P₂O₅ represented 0.23 percent of the total mass. Thus, not much phosphorus could have been in the vapor state, and none was found, even though P₂O₅ or H₃PO₄ are reasonably volatile compounds.

Sulfate in the fly ash from the ESP hoppers accounted for about 5 percent of the sulfur in the coal. On the other hand, SO₂ collected as sulfate found in the impinger solutions of the train for acid gases represented about 90 percent of the sulfur in the fuel.

Fluoride and chloride were not found in the fly ash. These halogens were collected in the train for acid gases at concentrations averaging 80 percent of the fluorine in the coal or 108 percent of the chlorine in the coal, presumably due to their occurrence as HF and HCl gases. These recoveries were for the ESP outlet; the recoveries were more variable and less complete at the inlet. Table 7 lists average concentrations of HF, HCl, and SO₂ for the flue gas at the ESP outlet based on the amounts of the elements collected in the impingers of the sampling train. The emission of the non-metals predominantly as gases was not influenced perceptibly either by low NOx combustion or operating with one or two ESPs.

DATA ON ORGANIC COMPOUNDS

Volatile Compounds

Volatile organic compounds were defined, effectively by the sampling and analytical methods used, as compounds boiling below 100 °C. Not all compounds thus defined could be collected and analyzed, however. One of the notable exceptions was formaldehyde, which was not detected

in samples from the volatil alternate procedure with a dit

The principal objective with determine the concentrations (GC/MS) method employed. The computer program used basis of chromatographic retamong the 189 compounds it testing a total of 19 identification.

A group of 16 PAHs were the primary targets for measurement. Only one of these compounds was detected with any significant frequency: naphthalene, which is the PAH with the most simple structure and lowest molecular weight. This compound occurred at concentrations near the limit of detection, around 10 μ g/Nm³ (note that this figure is three orders of magnitude below that sometimes seen for benzene).

Two compounds of lower molecular weight and higher volatility were seen consistently: phenol and benzoic acid (which is not on the list of 189 air toxics). The concentrations of these compounds were 100-600 µg/Nm³ at the ESP inlet and thus much higher than that of the PAHs. The concentrations were significantly less at the ESP outlet. As with volatiles, many unidentifiable chromatographic peaks were detected. The total emission of semi-volatiles was evidently increased by low NOx combustion, as expected, yet contrary to the effect seen with volatiles.

The emissions of semi-volatiles were possibly suppressed by the ESPs. This effect might be attributed to the presence of the compounds in the adsorbed state on precipitated ash except for the fact that no organic matter could be found on the ash recovered from the ESP hoppers. If the hopper ash analyses are correct, it is conceivable that oxidation by ozone in the corona regions of the ESPs removed organic compounds.

Aldehydes and Ketones

Two compounds were detected: formaldehyde and acetone. These are the most simple compounds in the two classes concerned. The concentrations at the ESP inlet ranged from 20-200 μ g/Nm³ for formaldehyde and from 1-20 μ g/Nm³ for acetone. The concentrations were consistently lower at the outlet. One logical explanation that can be offered for the apparent effect of the ESPs is that ozone oxidation occurred, just as may have occurred with the semi-volatile compounds.

No information on the effect of low NOx combustion on the emissions of these compounds was obtained because the samples for baseline operation were analyzed incorrectly and disposed of before the error was discovered.

CONCLUSIONS

Effects on Chemical Emissions Due to LNCFS Level III Conversion

The change from normal firing of pulverized coal to low NOx firing did not produce sharp changes in the emissions of elementary substances. This is hardly surprising for metals, which generally occur in the fly ash in oxidation states only problematically related to the conditions of oxidation in

the boiler. One exception occurred with a trace metal that can exist in different oxidation states; hexavalent chromium apparently was suppressed by low NOx firing. The absence of changes in emissions of the non-metals of concern as a consequence of low NOx firing is not surprising either; the principal forms of these elements are phosphate in the ash and HF, HCl, and SO₂ in the flue gas.

The shift to low NOx firing was expected to influence the emissions of organic compounds because the emissions of such substances are an effect of incomplete combustion. If elementary carbon is not burned completely, as seems to be the case, hydrocarbons and other organic compounds are not likely to be burned completely either. The effect of low NOx firing on these compounds, unfortunately, cannot be described simply or unambiguously from the results of this testing. The data appear to present the anomaly of opposing effects: reduced emissions of volatiles such as benzene and other simple aromatics, and increased emissions of semi-volatiles, possibly from unidentified compounds.

Control of Chemical Emissions by Hot-Side and Cold-Side ESP's

The hot-side and cold-side ESPs removed approximately 99.5 percent of the particulate material entrained in the flue gas at the ESP inlet. Since most of the trace metals were associated with the particulate phase, most of the trace metals were controlled by the ESPs. However, the ESPs failed particularly at controlling mercury and selenium, which were largely in the vapor state. From the point of view that the main control of total particulate matter occurred at 700 °F in the hot-side ESP, however, the control efficiency for most of the trace metals may be regarded as unexpectedly high.

The data suggest that some organic compounds were removed from the flue gas through the ESPs and air heater. This effect may have been due to the presence of the vapors on solids that were precipitated. It may have also been due in part to the oxidation of the vapors to undetected residues, because of the presence of the vigorous oxidant ozone in the corona regions within the ESPs.

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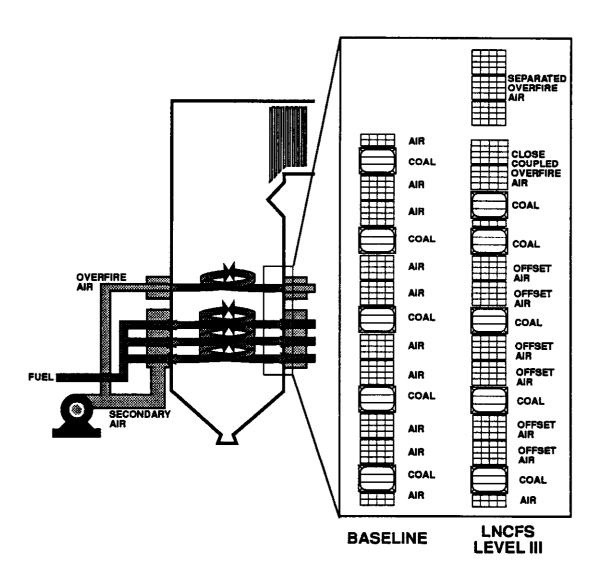


Figure 1. Low NOx Concentric Firing System Level III

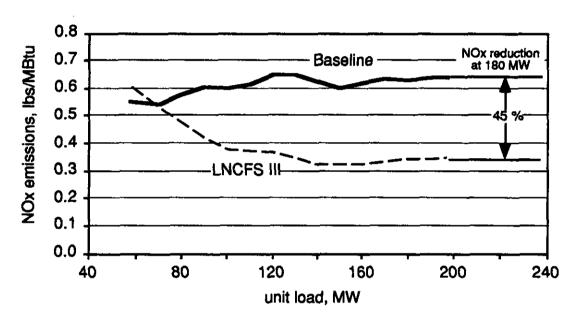


Figure 2. NOx Reduction with LNCFS Level III

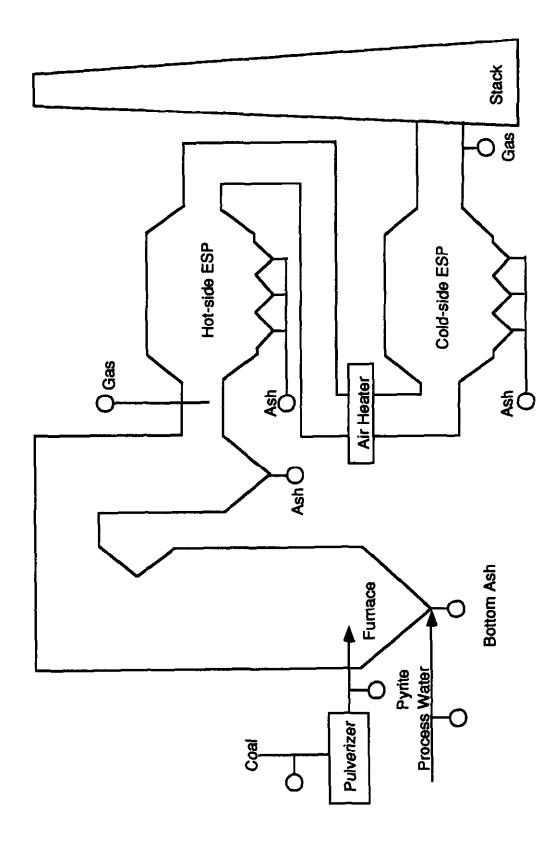


Figure 3. Layout of Sampling Locations

Table 1. PISCES Air Toxics List

Inorganic Chemicals	Polynuclear Aromatic Hydrocarbons (PAH) *	Polycyclic Organic Matter (POM) *	Volatile Organic Compounds
Arsenic	Acenaphthene	1,2-Diphenylhydrazine	Benzene
Barium	Acenaphthylene	1-Chloronaphthalene	Toluene
Beryllium	Anthracene	1-Naphthylamine	Formaldehyde
Cadmium	Benzo(a)anthracene	2-Chloronaphthalene	
Chlorine (Cl-)	Benzo(a)pyrene	2-Naphthylamine	
Chromium	Benzo(b)fluoranthene	3,3-Dichlorobenzidine	
Cobalt	Benzo(g,h,i)perylene	4-Aminobiphenyl	
Copper	Benzo(k)fluoranthene	4-Bromophenly phenyl ether	
Fluorine (F-)	Chrysene	4-Chlorophenyl phenyl ether	
Lead	Dibenzo(a,h)anthracene	Benzidine	
Manganese	Flouranthene	Butylbenzylphthalate	
Mercury	Fluorene	Dibenzofuran	
Molybdenum	Indeno(1,2,3-cd)pyrene	Dibenz(a,j)acridine	
Nickel	Naphthalene	Diphenylamine	
Phosphorus	Phenanthrene	n-Nitrosodiphenylamine	
Selenium	Pyrene		
Vanadium	2-Methylnaphthalene		
	3-Methylcholanthrene		
	7,12-Dimthyl- benzo(a)anthracene		

^{*} Semi-Volatile Organic Compounds

Table 2. Methods for sampling flue gas stream

Analyses	Sampling apparatus	Sampling media
Metals: As, Ba, Be, Cd, etc.	Method 5-type train known as the EPA Multiple Metals Train	Filter HNO ₃ -H ₂ O ₂ impingers H ₂ SO ₄ -KMnO ₄ impingers
Non-metals: F, Cl, S, P	Method 5-type train for "acid gases" or "anions"	Carbonate-bicarbonate- peroxide impingers
Volatile organic compounds	So-called "VOST" (Volatile Organics Sampling Train)	Tenax and charcoal absorbers; water condensate
Semi-volatile organic compounds	So-called "Modified Method 5 Train"	Filter XAD resin Water-filled impingers
Aldehydes, ketones	Method 5-type train with trapping compound DNPH	Impingers containing dinitrophenolhydrazine (DNPH)

Table 3. Analytical methods for solids and flue-gas constituents.

Analytes	Methods
Metals	
Ba, Be, Co, Cr, Cu, Mn, Mo, Ni, V	Inductively coupled argon plasma emission spectroscopy (ICP)
Cd, Pba	Graphite furnace AAS (GFAAS)
As, Se, Sb	Hydride generation AAS (HGAAS)
Hg	Cold-vapor AAS (CVAAS)
Non-metals	
S as sulfate	Ion chromatography
F as fluoride	Ion-specific electrode
Cl as chloride	Ion chromatography
P as phosphate	Ion chromatography or colorimetry
Volatile organics	Gas chromatography/mass spectroscopy (GC/MS)
Semi-volatile organics	Gas chromatography/mass spectroscopy (GC/MS)
Aldehydes, ketones	High performance liquid chromatography with UV detection

^aAnd others if required for sensitivity

Table 4. Illustrative data on mercury (baseline test at the ESP outlet, only hot-side ESP operating)

Form of mercury	Concentration, a µg/Nm ³
Particulate	Not determined
Vapor Hg(0) Hg(II) Methylmercury	1.85 7.37 0.0021
Total, all forms	9.22

^aData are averages from three sampling experiments.

Table 5. Recoveries^a of metallic elements in the gas stream entering the hot-side ESP

Element Antimony	Baseline testing Min. Max.,% Indeterminate ^C	Low NOx testing Min. Max., % Indeterminate ^C
Arsenic	81-120	Indeterminate ^C
Barium	69- 88	168-179
Beryllium	54-103	22-113
Cadmium	76-346	26-166
Chromium	128-173	67-112
Cobalt	64-145	Indeterminate ^C
Copper	47- 81	22- 37
Lead	131 -256	26- 46
Manganese	91-121	64- 90
Mercuryb	100-134	70-106
Molybdenum	97-179	84-105
Nickel	86-124	86-121
Selenium	59- 61	49- 98
Vanadium	55- 74	56- 61

a Recovery is the percentage of element in the coal found in the gas stream. The data are from three tests in each series with the cold-side ESP de energized.

b All data except for mercury are for the ESP inlet; the data for this element are from the outlet and are believed correct for the inlet since a negligible fraction of this element was in the particulate state.

c Results that are shown as indeterminate can be illustrated in this way: The concentration of antimony in the coal has to be reported as giving a flux of <5.0 g/min, a value consistent with the value based on the coal but still not providing a figure for recovery.

Table 6. Material balance across the ESP system.

Element	Closure, a %, Baseline testing	Closure, ^a %, Low NOx testing
Antimony	<40	95
Arsenic	132	161
Barium	60	200
Beryllium	>69	36
Cadmium	95	211
Chromium	131	134
Cobalt	117	133
Copper	109	104
Lead	99	138
Manganese	114	123
Mercury	<57	<30
Molybdenum	107	89
Nickel	117	102
Selenium	Indeterminate	5
Vanadium	111	123

Closure is the percentage of the element removed from the gas stream that is found in the hopper ash. The data on removal are based on inlet and outlet concentrations plus flow rate. The data on hopper accumulation rate are based on the solids analysis plus the amount of entrained solids that is collected in the hot-side ESP. The data given here are averages for four tests in each series, one test with both ESPs operating and three tests with only the hot-side unit operating.

Table 7. Concentrations of acid gases and corresponding recoveries of non-metallic elements in the coal.

Gas	Concentration, ppm	
HF	6.7	
HCl	117	
SO_2	2080	

Table 8. Concentrations of benzene and toluene.

Test series	Sampling time, min	Benzene		Toluene	
		ESP Inlet	ESP Outlet	ESP Inlet	ESP Outlet
Baseline	40		500±200		20±20
	10	2800±1300	1980± 20	310±320	13±14
	2	10,000	3500±1500	4300	50±67
Low NOx	10	160±220	1090± 430	7.1±2.6	6.6±3.4
	5	230±280	1200± 590	2.4±2.3	5.7±3.0
	2	310±260	1850±1930	3.0±6.0	6.7±7.1

Data are averages and standard deviations except for sampling times that yielded on single results.

Results of Babcock & Wilcox's Clean Coal Technology Combustion Modification Projects: Coal Reburning for Cyclone Boiler NO_x Control and Low NO_x Cell™ Burner Demonstrations

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- (1) Four B&W reburn burners
- (2) Four standard dual air zone overfire air ports
- (3) An MPS-67 pulverizer and primary air fan
- (4) 150 ton coal silo
- (5) Pulverizer enclosure building
- (6) Control system modifications
- (7) Reburn motor control center and power supply transformer
- (8) Various flues, ducts, flow control dampers and monitors

The isometric view of the system shown in Figure 3 gives the spacial relationships of each of the components in the system. Integration of the reburn system with the existing plant consists of interfaces with the coal feed tripper conveyor, the air heater outlet, flue gas recirculation system, forced draft fan discharge, hot air recirculation system, penetrations into the boiler, and the control system. Tie-in of all reburn components was accomplished during the Fall outage, from September 16 through October 31, 1991.

Coal Reburning Test Results

The primary test coal for the coal reburning demonstration was an Illinois Basin bituminous coal (Lamar). The majority of the testing was performed while firing this fuel to reflect the higher sulfur bituminous coal fired by many of the utilities operating cyclones. Following the bituminous coal testing, subbituminous Powder River Basin (PRB) coal tests were performed to evaluate the effect of coal switching on reburn operation. In addition, WP&L's strategy to meet sulfur emission limitations as of January 1, 1993 is to fire the low sulfur coal.

Reburning Test Parameters

There were three sequences of testing of the coal reburning system using Lamar coal. Parametric optimization testing was used to set up the automatic controls. Performance testing was run with the unit in full automatic control at set load points. Long-term testing was performed with reburn in operation while the unit followed system load demand requirements. PRB coal was tested by parametric optimization and performance modes.

A test matrix was established in order to determine optimized operation. The test variables included in the matrix along with the approximate ranges tested are:

- Boiler load (37 to 118 MW_c)
- Reburn system percent of total boiler heat input (≈25 to 40%)
- Reburn zone stoichiometry (≈0.83 to 0.96)
- Reburn burner stoichiometry (≈0.35 to 0.70)

- Reburn burner pulverized coal fineness (80 to 98% through 200 mesh)
- Gas recirculation rates to reburn burners (0 to 5% of boiler)
- · Reburn burner spin vane and impeller/swirler adjustments
- · Overfire air (OFA) port spin vane/sliding disk adjustments
- Economizer outlet 0,% (2 to 4%)

NO, and CO Emissions

Baseline (no reburning) data for NO_x emissions under various load conditions for both coals are summarized in Figure 4 and in Table 2.

TABLE 2 - Baseline NO, Levels for Lamar and PRB Coals				
Baseline NO _x Emissions - ppm (lb/10 ⁶ Bt Corrected to 3% Oxygen				
Load (MW _e)	Lamar Coal	Powder River Basin Coal		
118	635 (0.86)	-		
110	609 (0.83)	560 (0.75)		
82	531 (0.72)	480 (0.64)		
60	506 (0.69)	464 (0.62)		
38	600 (0.82)	-		

 NO_x levels increase at 38 MW, during Lamar firing because the boiler goes to single cyclone operation, approaching the heat release conditions and corresponding NO_x emissions achieved at full load.

CO emission levels during baseline operation were low while firing either of the two coal types. Generally speaking, the CO levels were slightly lower during the PRB coal firing tests (approximately 30 to 45 ppm versus 60 to 70 ppm over the load range).

Reburn testing on both the Lamar and PRB coals indicates that varying reburn zone stoichiometry is the most critical factor in changing NO_x emission levels during coal reburning operation. The reburn zone stoichiometry can be varied by altering the air flow quantities (oxygen availability) to the reburn burners, the percent reburn heat input, the gas recirculation flow rate or the cyclone stoichiometry.

Figure 5 represents B&W economizer outlet NO_x and CO emission levels in ppm corrected to 3% O_2 versus reburn zone stoichiometry at full load conditions (110 MW_c) while firing Lamar coal. This figure consists of parametric optimization and performance testing data. Figure 6 presents NO_x and CO emissions while firing PRB coal.

Load versus NO_x emissions for both coals are shown in Figure 7 and summarized in Table 3.

TABLE 3 - Reburn NO _x Emissions Versus Load for Lamar and PRB Coals			
T - 4 (10m)	Reburn NO _x Emissions/% Reduction from Baseline (ppm/%)		
Load (MW _e)	Lamar Coal	PRB Coal	
118	-	275/-	
110	290/52	208/62	
82	285/47	215/55	
60	325/36	220/53	
41	<u> </u>	220/-	

Reburn operation burning PRB produced lower overall NO_x emission levels. Baseline NO_x levels with PRB were approximately 10% lower, and better NO_x reduction is probably due to the higher Western fuel volatile content. Higher volatile content generates higher concentrations of hydrocarbon radicals in the substoichiometric region of the furnace. Figure 7 also shows that PRB NO_x emissions could be maintained at a constant level over the 110 to 41 MW_c load range.

With PRB coal, at loads higher than 110 MW_c, NO_x emissions increased. At 118 MW_c, the NO_x level was 275 ppm (0.37 lb/ 10^6 Btu). Higher NO_x was due to less percent reburn heat input because of reburn feeder limitations. No baseline NO_x levels were obtained at this higher load because the boiler could not reach it without reburn burners in service.

Electrostatic Precipitator Performance

Considerable analysis was conducted on precipitator parameters during the initial stages of the project. It was anticipated with the Lamar coal that particulate loading would increase by as much as two times, depending upon the percentage of reburn fuel used. The analysis suggested that stack opacity would increase to 18 to 20% (the unit has a 40% opacity limit).

When the reburn system was operated, the opacity remained unchanged or decreased slightly. The results of several precipitator tests showed that the particulate grainloading to the precipitator increased about 37%, much less than the two times expected, while outlet grainloading decreased slightly. In general, precipitator efficiency increased slightly with reburn operation. This is probably the result of increased flyash mean particle size (43% of baseline particles were less than 2 microns in size versus 27% with reburn) and no change in flyash resistivity, which offset increased precipitator inlet grain loading.

The precipitator performance did not change significantly with PRB coal. Opacity was consistent with Lamar coal tests. Increases in inlet grain loading (with the reburn system in service) were not as great as that seen with Lamar coal (20% or less versus 30%). Outlet grain loading and precipitator efficiency were generally unchanged from baseline conditions. There was no apparent change in the flyash to total ash ratio.

Unburned Carbon Efficiency Loss

Figure 8 is a plot of change in unburned carbon boiler efficiency loss (UBCL) from baseline conditions versus steam flow (an indication of boiler load) for both Lamar and PRB coals with reburn in operation. For Lamar coal, the full, medium and low loads UBCL were 0.1, 0.25 and 1.5% higher, respectively, than baseline. Full, medium and low load UBCL increases with the PRB coal during reburn operation were 0.0, 0.2 and 0.3%, respectively. Combustion efficiency improved with PRB fuel as did reburn burner flame stability.

Furnace Exit Gas Temperature

Figure 9 shows the FEGT with and without reburn in service for the two coals tested. At full load firing the Lamar coal, the FEGT decreased by approximately 100 to 150F with reburn in service. The gas recirculation flow with reburn in service would be expected to cause about 25F of this decrease. There was no change in FEGT at 75% load and an increase of 50 to 75F at 50% load with reburn in service.

For the PRB coal tests at full load, the FEGT decreased by approximately 25 to 50F with reburn in service. Once again, the gas recirculation flow with reburn in service would account for approximately 25F of this change. There was no change in FEGT at 75% load and an increase of 75F at 50% load with reburn in service. The FEGT decreases at full load in both cases were reflected in significantly decreased superheater and reheater attemperator spray flows.

Although the explanation for this phenomenon is still unclear, it

is believed that changes in emissivity in the furnace under substoichiometry conditions is causing increased furnace heat absorption.

Slagging and Fouling

During reburn system operation with Lamar coal, the operators continually monitored both the boiler internals for increased ash deposition and the On-Line Performance Monitoring System (OPM) for heat transfer changes. At no time throughout the system optimization or long term operation period were any slagging or fouling problems observed. In fact, during the scheduled spring and fall unit outages, internal boiler inspections revealed that boiler cleanliness had actually improved.

Because slagging and fouling is usually time dependent, experience on PRB coal is limited. OPM monitoring of furnace and convective pass heat transfer surfaces indicated no change over baseline, Lamar coal conditions. This is an improvement over previous PRB coal experience (without the reburn system) where careful monitoring of slagging and fouling conditions was required. PRB coal will be burned in the unit in the future and additional information and experience will be gained.

Furnace Corrosion

During the major reburn system installation outage (Fall 1991), extensive furnace wall tube ultrasonic thickness (UT) measurements were taken. In Fall 1992, at the completion of the long term testing, and again during the next scheduled outage in Spring 1993, UT measurements were taken in the same areas of the furnace. Additionally, tube specimens were removed from the rear wall of the furnace in the reburn zone for destructive examination. No observable decrease in tube wall thickness was measured. Follow up UT testing will continue for the next five years.

Hazardous Air Pollutant (HAP) Testing

Hazardous air pollutant (HAP) testing was performed using the Lamar test coal at the request of DOE and EPRI to assess the technology's environmental performance. The work was performed near the end of the testing program. The following streams were sampled:

- Crushed coal from the cyclone feeders
- Reburn coal from the pulverizer outlet
- Molten slag from the furnace
- Flue gas at the precipitator inlet
- Flue gas at the precipitator outlet
- Flyash from the precipitator hoppers

The trace elements analyzed were arsenic, beryllium, cadmium, chromium, lead, nickel, manganese, selenium and mercury. Volatile and semivolatile organics (benzene and toluene), aldehydes and acid gases (hydrogen fluoride and hydrogen chloride) were also tested.

HAP emissions were generally well within expected levels and emissions with reburn were comparable to baseline operation. No major effect of reburning on trace metals partitioning was discernable.

None of the 16 targeted (by Title III of the 1990 CAAA) polynuclear aromatic semivolatile organics were present in detectable concentration, at a detection limit of 1.2 ppb for either baseline or reburn operation. Of the 28 targeted volatile organics analyzed, the only compounds present at detectable levels were benzene and toluene and these are summarized in Table 4.

TABLE 4 Hazardous Air Pollutant Emission Results for Cyclone-Fired Boilers-Organics				
Test Condition	Toluene, ppb	Benzene, ppb	Semivolatile PNA, ppb	
Average Baseline	0.38	0.84	<1.19	
Average w/Reburn	0.44	0.25	<1.60	

Aldehydes were not detectable at the 2.8 ppb level for formaldehyde and 1.9 ppb level for acetaldehyde.

Reburn Results Summary

Table 5 presents a comparison of anticipated and actual results of reburn operation. The reburn system has performed very well as evidenced by WP&L's decision to continue system operations beyond the term of the DOE Coal Reburning Project.

A significant advantage of coal reburning is that it minimizes and possibly eliminates a 10 to 25% derate normally associated with switching to a PRB coal in a cyclone unit. The derate is a result of using of lower Btu content fuel in the volume limited cyclone. The reburn system transfers about 30% of the heat input out of the cyclones to the reburn burners, bringing the cyclone feed rate down to a manageable level, while maintaining full load heat input to the unit. At Nelson Dewey, maximum pre-reburn retrofit full load on PRB coal was 108 to 110 MW, while on the

higher Btu Lamar coal, 118 MW, could be achieved. With reburn in operation, the unit was able to achieve 118 MW, on PRB coal. Accordingly, a reburn system possibly could be economically justified based on fuel cost savings and regained unit capacity when switching to a PRB coal.

TABLE 5 Effect of Reburn System on Unit Performance			
Parameter	Anticipated Results	Actual Results	
NO _x emissions (full load) Illinois Basin coal	Reduced 50% or more	Nominal 55% reduction	
NO _x emissions (full load) Powder River Basin coal	Reduced 50% or more	Nominal 61% reduction	
Precipitator opacity	Up 5 to 10%	No increase from base	
Slagging/fouling	No change	Cleaner than normal	
Furnace corrosion	No change	No change	
Header/tube temps.	Higher 25 to 50F	No increase from base	
FEGT (Illinois Basin - Lamar coal)	Higher by 50 to 75F	Reduced by 100 to 150F	
FEGT (PRB)	Higher by 50 to 75F	Reduced by 25 to 50F	
SH & RH sprays (Illinois Basin - Lamar Coal)	Higher by 30%	50% of base	
Unburned carbon efficiency loss (full load) Illinois Basin coal	Higher	Higher by 0.1%	
Unburned carbon efficiency loss (full load) Powder River Basin coal	Higher	No change	
Hazardous air pollutants (Illinois Basin - Lamar coal)	No change	No change	

LOW NOx CELL BURNERS (LNCB)

Description of Technology

The original cell burner design consisted of two or three circular burners mounted in the lower furnace. Figure 10 shows a two-nozzle cell burner. The two-nozzle LNCB shown in Figure 11 was developed by B&W in association with the EPRI. The features of the LNCB were designed to minimize the formation of thermal and fuel NO. The two original circular burners in each cell are replaced with a single S-type circular burner and a close coupled secondary air injection port. The flame shape is controlled using an impeller at the exit of the burner and adjustable spin vanes in the secondary air zone. The air port louver dampers provide additional control over the mixing between the fuel and The S-burner operates at a low air-fuel air streams. stoichiometry, typically 0.6, with the balance of air entering through the adjacent air port. The delayed mixing of the fuel and air during the initial stage of combustion limits the formation of NO..

Low NO. Cell Burners at J. M. Stuart Station Unit No. 4

The host site for the full scale demonstration of the LNCB was DP&L's J. M. Stuart Station Unit No. 4 (JMSS4). JMSS4 is a B&W 605 MW. Universal Pressure (UP) boiler, a once-through design, originally equipped with 24, two-nozzle cell burners arranged in an opposed wall configuration as shown in Figure 12.

Each of the original two-nozzle cell burners were replaced with a single S-type circular burner in place of the lower cell burner and a close coupled secondary air injection port at the upper cell location, shown in Figure 11. To avoid replacing coal pipes and pulverizer top housings, the two coal pipes, one to each burner of the original cell, were combined at the burner front to supply the new single S-type circular burner by using a special Y-pipe assembly. As a special feature of the LNCB technology, no pressure part modifications were necessary and the existing control system was utilized. The retrofit of the LNCB equipment was completed during a six week scheduled turbine outage during October/November 1991.

Initial test results with this original arrangement (Figure 13) indicated high levels of CO and hydrogen sulfide (H_2S) in the lower hopper region of the furnace, an unacceptable operating condition in this pressurized furnace. As a demonstration project, resources were allocated to perform in depth background work to develop the numerical model to help understand flow behavior in the unit. When problems with the LNCB operation arose, B&W used its three dimensional numerical modeling capabilities to simulate the existing operating condition, as

well as evaluate alternative burner/secondary air port arrangements that could mitigate this problem. The best computer generated analysis identified for maximum mitigation of CO and H₂S levels was to invert the air port and burner of every other LNCB on the lowest level of burners (Figure 14). This is the final configuration for which results are subsequently reported in this paper.

A second result of initial testing showed that NO_x reduction of only 35% from baseline levels was being achieved with the 50 degree coal impellers. By retracting the impellers within the coal nozzles, NO_x reduction increased to 45%. This indicated a need for an impeller design change in order to achieve the NO_x reduction goals of the project. A coal impeller with a 25 degree included angle was designed, fabricated and installed during the same one week outage in April 1992 in which the alternating inverted LNCB arrangement was accomplished.

Low NO Cell Burner (LNCBTM) Test Results

The LNCB demonstration emphasized evaluation of boiler performance, boiler life and environmental impact. Key boiler performance parameters that were measured included boiler output (steam temperatures); flue gas temperatures at the furnace, economizer and air heater exits; the slagging tendencies of the unit; and UBC losses. Evaluation of $\rm H_2S$ levels, ultrasonic testing of lower furnace tube wall thicknesses and destructive examination of a corrosion test panel were the mechanisms used to predict impact on remaining boiler life. Environmentally, $\rm NO_x$, $\rm CO$, carbon dioxide ($\rm CO_2$), total hydrocarbons (THC) and particulate matter, dust loadings and precipitator collection efficiency were measured at varying test conditions.

NO., CO Emissions and Unburned Carbon Losses

Full Load, 6 Mills In Service (Avg. 604 MW.)

At full load conditions, averaging 604 MW, with all mills in service, average NO_x emissions were 0.53 lb/10⁶ Btu of heat input to the unit. This represents a NO_x reduction of 54.4%, averaging all data. Figure 15 presents NO_x data for both baseline (preretrofit) and post-retrofit operation as a function of excess air.

Emissions of CO under the same conditions ranged from 28 to 55 ppm.

The weighted average of unburned carbon content in ash (UBC) for samples collected from the boiler bottom ash hopper, the boiler outlet hopper and the precipitator first field hopper was 1.12% during full load operation, all mills in service, averaging 604

MW_c. This represents an unburned carbon efficiency loss (UBCL) of 0.2%. This is a 56% improvement over baseline unburned carbon losses and is most likely the result of improved air flow distribution provided by the LNCB retrofit.

Full Load, 5 Mills In Service (Avg. 604 MW.)

A total of six tests were conducted at full load with a different mill out of service for each test. Each mill provides pulverized coal to four LNCBs. Figure 16 shows burner/mill combinations.

The average NO_x emissions level for full load, five mills in service was approximately 0.51 lb/10⁶ Btu. This represents an average reduction from baseline conditions of 53%. Figure 17 presents the NO_x data for both baseline and post-retrofit conditions.

NO_x emissions were lowest at approximately 0.48 lb/10⁶ Btu when either of mills A or F was the out-of-service mill. These mills fire the upper outer two burners on each side of the furnace. The highest NO_x levels occurred when mill D was out of service, at 0.56 lb/10⁶ Btu. Mill D fired the lower outer two burners on each side of the front wall. Mill C out of service also experienced higher NO_x emissions at 0.52 lb/10⁶ Btu as the lower outer two burners on each side of the rear wall were taken out of service. Apparently, with upper burners out of service and the remaining burners firing harder, slightly more NO_x reduction is achieved, possibly due to deeper staging of the lower burners followed by more secondary air available at the burner out-of-service level.

The average CO emissions rate ranged between 20 and 38 ppm during one mill out of service testing. The weighted average for UBC samples was 2.52%. This represents a small reduction from baseline UBC levels which translates to a small improvement in UBC efficiency losses from 0.46% baseline to 0.42% post-retrofit.

Intermediate Load, 5 Mills In Service (Avg. 460 MW.)

For these tests, Mill A was chosen to be out of service because NO_x emissions at full load A mill out of service were among the lowest observed with one mill down. Figure 18 shows NO_x emissions versus excess air for this test mode. Average NO_x emissions rate for intermediate load condition with five mills running was 0.42 lb/10⁶ Btu input corresponding to a 54% reduction in NO_x emissions from similar baseline conditions.

The average CO emissions rate for this intermediate load condition ranged between 28 and 45 ppm. The weighted average of UBC for all sample locations averaged 0.98% for all of the tests at this condition. The efficiency impact due to unburned

combustibles loss is 0.17%. This reflects a decrease in the carbon-in-ash levels from those obtained during the baseline tests and also represents a 64% improvement in UBC efficiency losses when compared with the baseline case.

Low Load, 4 Mills In Service (Avg. 350 MW.)

For low load conditions, mills A and F were out of service because best NO_x reduction was achieved at full load with the upper burners out of service. They were also chosen to test the ability of the LNCB's to maintain low NO_x while the boiler was pushed to maintain reheat superheater steam temperature. This condition represents the original reheat superheater outlet temperature control point.

Figure 19 shows NO_x emissions rate versus excess air for baseline and post-retrofit test conditions at low load. The average NO_x level was 0.37 lb/10⁶ Btu which represents a reduction of about 48% from baseline. Emissions of CO ranged from 5 to 27 ppm. The weighted average of UBC for all sample locations averaged 3.17% for all tests, which represents a 0.59% efficiency loss due to unburned carbon. This is an 18% increase in efficiency loss compared to baseline results.

One-Day Test

On March 1, 1993, one day of emissions testing was conducted with all mills in service at JMSS4. The purpose of the test was to evaluate NO, emissions along with flyash UBC levels eight months after completion of optimized testing. The results are shown in Figure 20. No. averaged 362 ppm (0.49 lb/106 Btu) at 2.6% 0, (dry) at an average boiler load of 603.5 MW. The fuel used during the test averaged 11,736 Btu/lb with 14.1% ash and a fixed carbon to volatile matter (FC/VM) ratio of 1.45. Flyash grab samples obtained from the first field of the precipitator hoppers and bottom ash samples were analyzed for UBC. The weighted average UBC for the one-day test was 0.97%. This is a very good result, however, it is based on grab samples of ash in the precipitator hoppers and was not isokinetically collected. Basically, this shows no problem with UBC, which was the purpose of the grab samples.

Long Term Averages

An important aspect of the project was to record NO_x emission levels from JMSS4 during normal load dispatch operations over a long period. Table 6 shows the average NO_x emissions for JMSS4 with all mills in service as recorded by the Acurex CEM equipment through a total of two probes located one in each of the east and west economizer outlet ducts. This data was acquired between August 1992 and March 1993 during periods when the boiler was

operating above 590 MW_c. The number of days in each month with all mills in service, full load conditions is shown in the first column. All other days represent operation at lower load; mills out of service; or the Continuous Emissions Monitor (CEM) out of service or in calibration. The average NO_x level achieved for the eight month period was 0.49 lb/10⁶ Btu or a 58% reduction from baseline. The highest monthly average NO_x level observed was in January at 0.56 lb/10⁶ Btu. Wet coal and accompanying problems were suspected to have caused the higher level which still represented a 52% reduction. The excess O₂ levels averaged 3.2%.

TABLE 6 - Long Term Full Load All Mills In Service Data All Mills In Service Averages at JMSS4 Acurex CEM Test Results for Loads Above 590 MW.								
Month	Days	All Mills In Service						
		Load MW.	Dry O ₂ Econ Out	Dry NO _x ppm Corr. to 3% O ₂	NO. lb/10 Btu			
August	8.54	604	3.7	367	0.50			
September	7.29	604	3.2	333	0.45			
October_	14.51	605	3.3	367	0.50			
November	12.03	605	3.2	345	0.47			
December	4.94	605	3.1	360	0.49			
January	6.83	605	3.2	410	0.56			
February	7.22	606	3.2	364	0.50			
March	17.66	602	2.9	353	0.48			
Weighted 8-month avg.		604	3.2	360	0.49			
Total Days	79.02							

Table 7 shows the full load, mill out of service NO_x emission levels recorded during this same period. The lower NO_x levels recorded with either A or F mill out of service, as observed previously, can be attributed to the fact that these mills feed the burners on the upper elevation only.

Overall unit efficiency remained essentially unchanged from baseline to optimized LNCB operation. The current operation of

JMSS4 at a lower overall excess air since optimization, has reduced the dry gas loss and increased boiler efficiency slightly.

TABLE 7 - Long Term Full Load Mill Out of Service Data								
Mill Out of Service Averages at JMSS4 Acurex CEM Test Results for Loads Above 590 MW,								
Mill Out of Service	Days	August '92 - March '93						
		Load MW.	Dry O ₂ Econ Out	Dry NO _x ppm Corr. to 3% O ₂	NO. lb/10° Btu			
A	1.04	603	3.4	314	0.43			
В	1.81	608	3.6	361	0.49			
С	1.41	602	3.5	388	0.53			
D	2.29	602	3.6	404	0.55			
E	3.02	606	3.3	357	0.49			
F	8.48	604	3.9	314	0.43			
Weighted 8-month avg.		604	3.7	343	0.47			
Total Days	18.05							

Corrosion Studies

During burner installation in October/November 1991, a corrosion test panel was installed on the boiler side wall between the upper and lower burner rows to evaluate corrosion potential. The panel consists of SA-213T2 bare tube material, aluminized spray coated T2 tube material, 309 L and 308 L stainless weld overlays on T2 tube material and a chromized T2 tube material. In addition, UT measurements were conducted in the furnace.

Preliminary analysis from destructively examining the furnace wall samples taken from the corrosion test panel show localized corrosion near the center of the panel. Tube thickness wastage readings on the bare T2 material ranged from as little as 0.002 in. (2 mils) to a maximum of 0.015 in. (15 mils) per year for the 15 months of operation. This 15 months also includes the 6 months of operation prior to the burner inversion when high levels of CO and H_2S were present in the lower furnace. The amount of wastage also varied with the tube metal temperature,

i.e., second pass tubes experienced slightly higher losses than did first pass tubes. These wastage rates are not significantly higher than those experienced on the side walls in the burner zone with the original cell burners in place.

The coated tubes in the corrosion test panel experienced no loss (wastage) of materials. Analysis of the bare T2 material above the burner zone, below the burner zone and around the burners also indicated no metal loss.

UT testing of the furnace will continue over the next five years to evaluate corrosion potential.

CONCLUSIONS

Both the Coal Reburning and LNCB projects have achieved the respective Clean Coal Program objectives. Both technologies have demonstrated NO, reductions in excess of 50% without significant adverse impact to other boiler emissions streams. The host site units have each continued to reach pre-retrofit full load output without significant impact to boiler operation. Results of long term emissions testing indicate performance has continued to exceed the project goals for each technology and both DP&L and WP&L⁸ have decided to operate the respective Clean Coal Technologies beyond the project end dates.

The low cost and short outage time for a LNCB retrofit make the design financially attractive. In a typical retrofit installation, the capital cost will include the LNCB hardware, coal pipe modifications, hangers, support steel, sliding air damper drives and associated electrical, with a capital cost of about \$5.5 to \$8.0 per kW in 1993 dollars, based upon the DOE 500 MW, reference unit for material and erection. The outage time can be as short as five weeks because the LNCB is a plug-in design.

For cyclones, coal reburning offers a NO_x reduction alternative at a higher price. Costs are expected to be in the \$65/kW range for a 100 MW_e unit and in the \$40/kW range for a larger 600 MW_e unit. Unlike a burner retrofit which already has coal handling and pulverizers/coal piping in place, this equipment must be included in the cost of a reburn system. Site specific factors related to pulverizer location and coal supply can greatly influence overall reburn system cost. However, coal reburning brings with it benefits allowing increased flexibility in coal selection which can yield significant fuel savings.

Corrosion potential will continue to be investigated over the next five years for both technologies.

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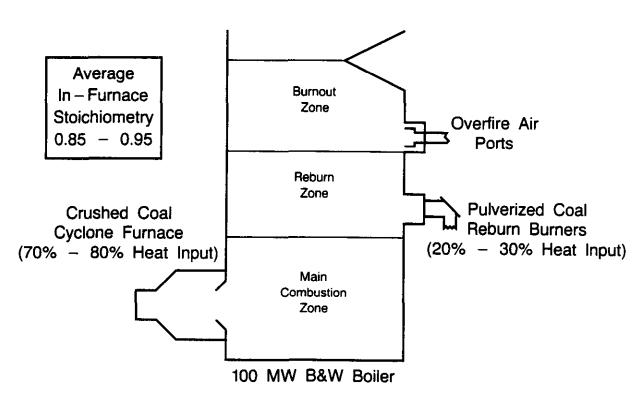


Fig. 1 Cyclone reburn combustion zones.

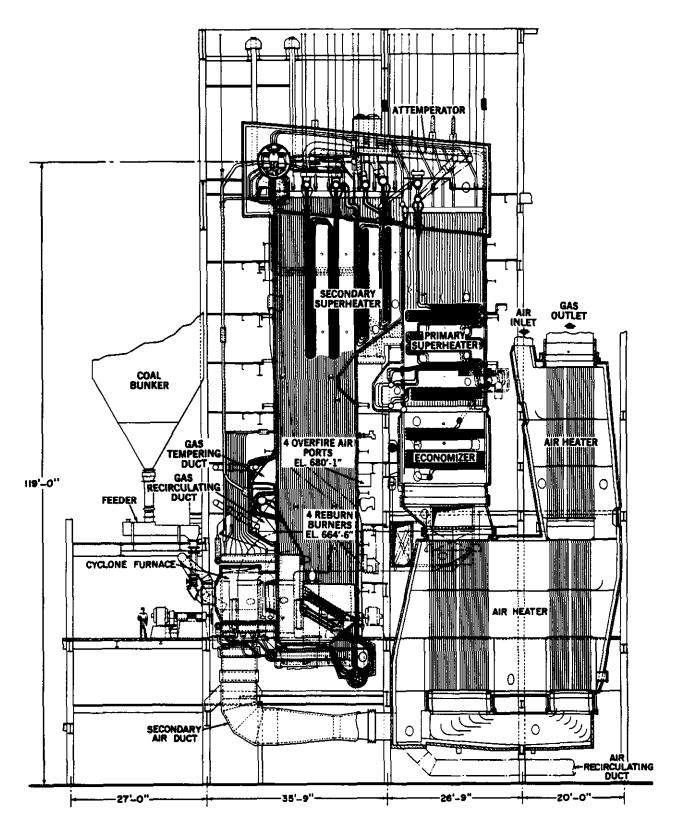


Fig. 2 WP&L Nelson Dewey Unit No. 2.

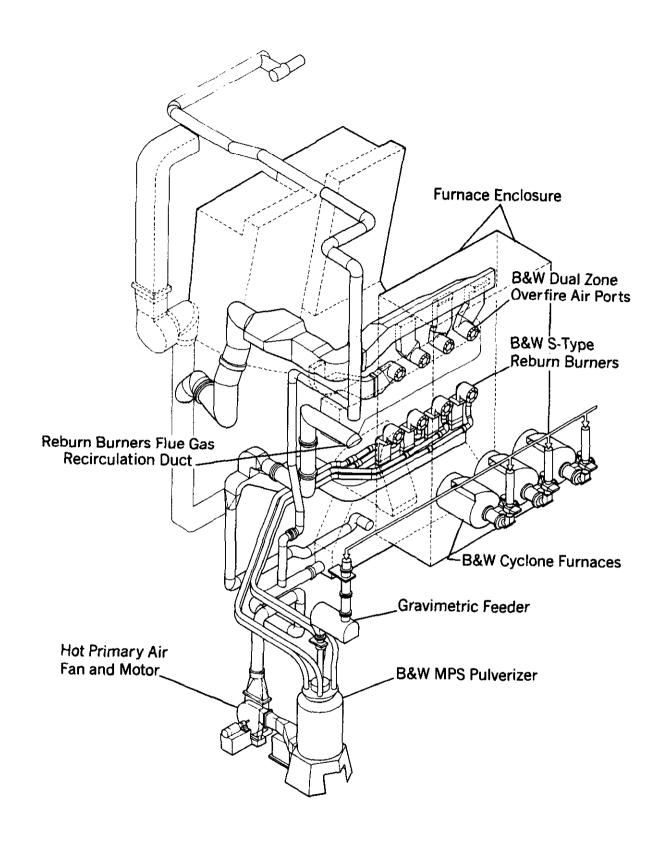


Fig. 3 Isometric view of Coal Reburning for Cyclone Boiler NO_x Control.

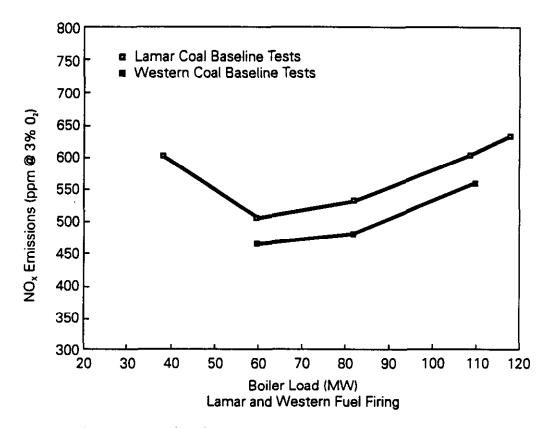


Fig. 4 Baseline NO_x emissions versus load, Nelson Dewey Unit 2.

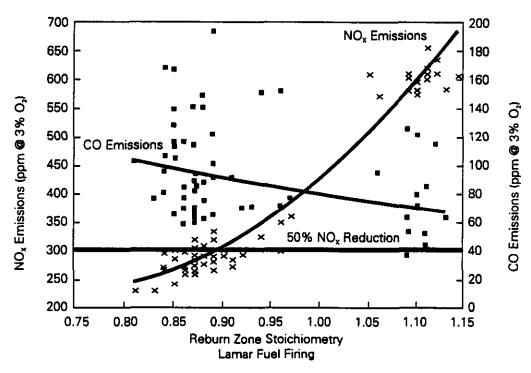


Fig. 5 NO, and CO emissions versus reburn zone stoichiometry at full load firing Lamar coal.

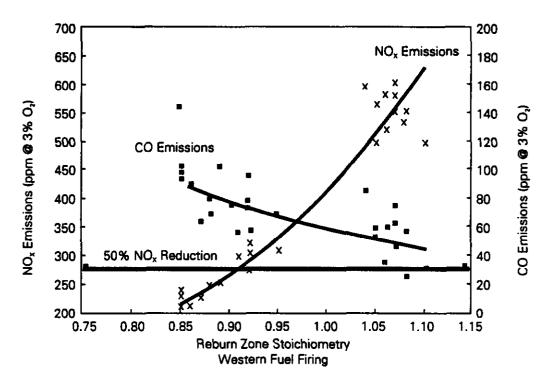


Fig. 6 NO_x and CO emissions versus reburn zone stoichiometry at full load firing Western coal.

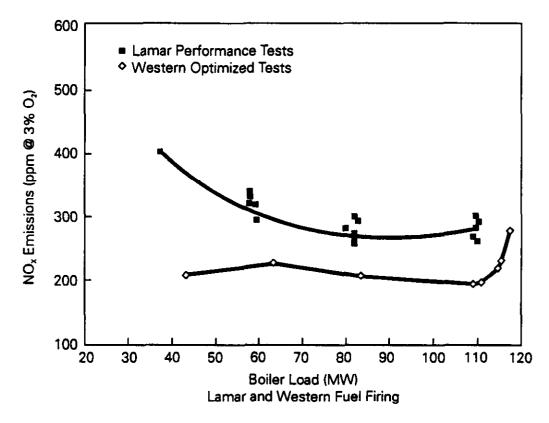


Fig. 7 NO_x emissions versus load firing Lamar and Western coals with reburn system in operation.

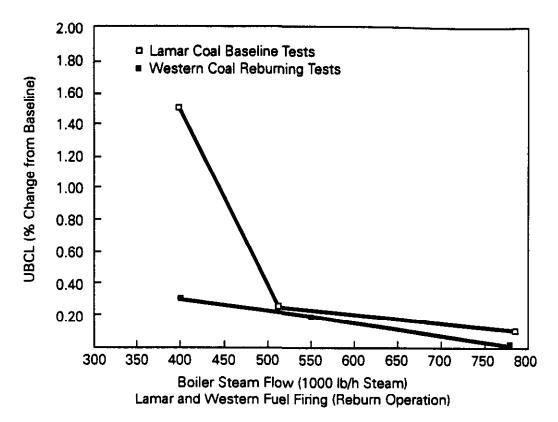


Fig. 8 Unburned carbon efficiency loss versus load with reburn system in operation.

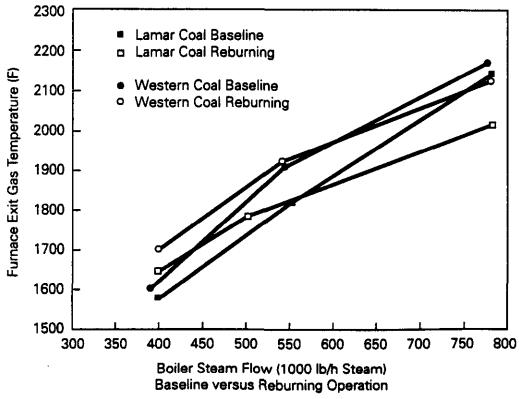


Fig. 9 Comparison of baseline and reburn system furnace exit gas temperatures.

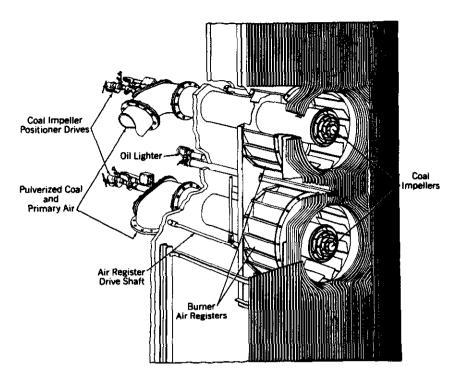


Fig. 10 Standard two-nozzle cell burner.

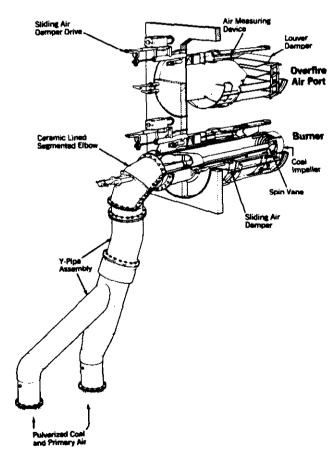


Fig. 11 Low NO_x Cell^m burner.

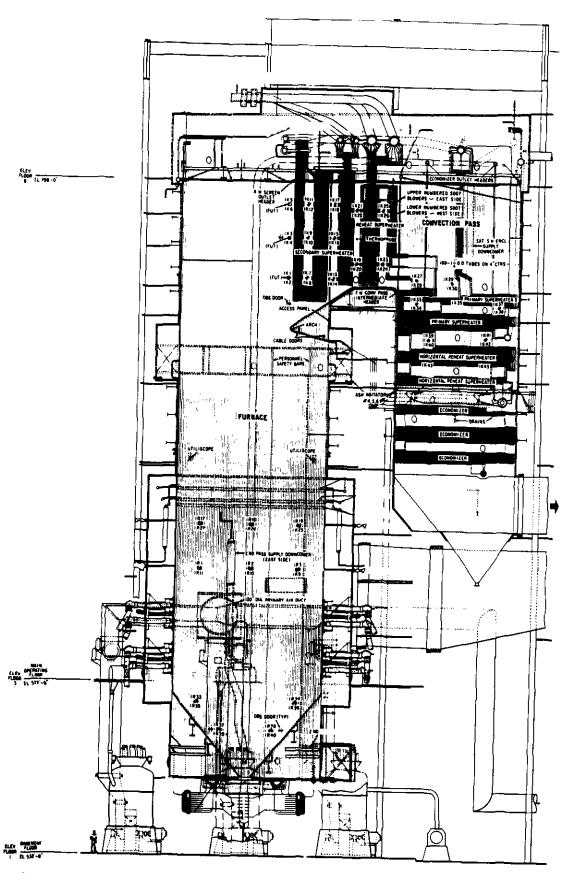


Fig. 12 DP&L J.M. Stuart Station Unit No. 4.

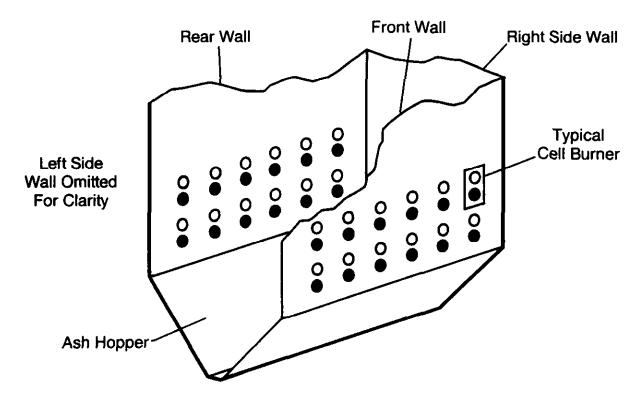


Fig. 13 Original LNCB™ arrangement.

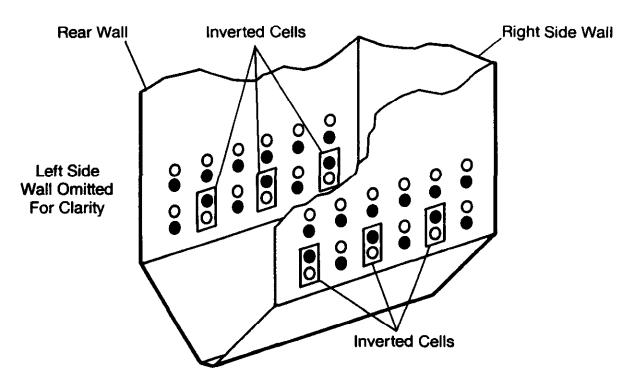


Fig. 14 Partially inverted LNCB™ arrangement.

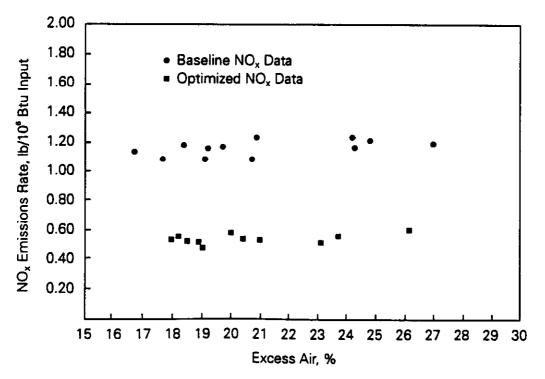


Fig. 15 LNCB $^{\text{m}}$ baseline and retrofit NO $_{x}$ emissions data as a function of excess air.

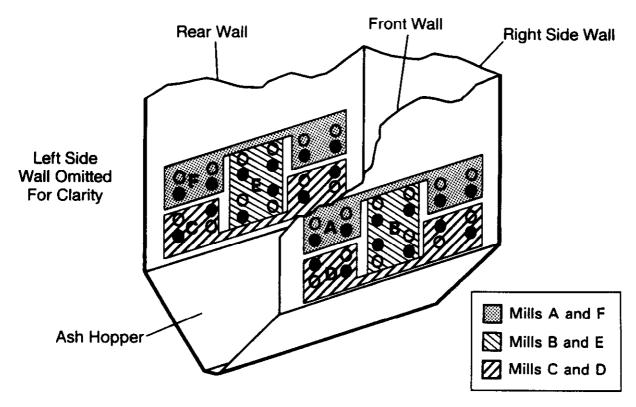


Fig. 16 Revised LNCB™ arrangement with burner/mill combinations.

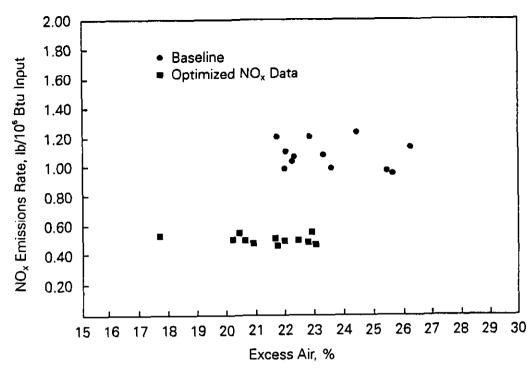


Fig. 17 LNCB™ NO_x emissions data for full load, five mills in service.

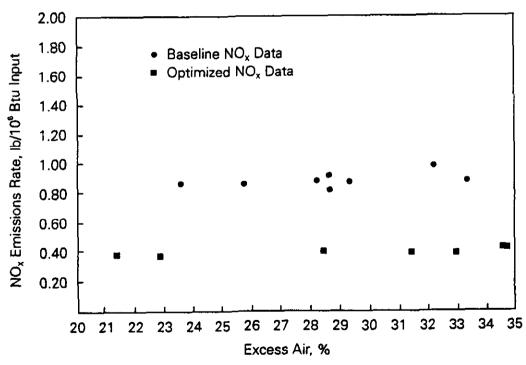


Fig. 18 LNCB™ NO, emissions data for intermediate (75%) load, 5 mills in service.

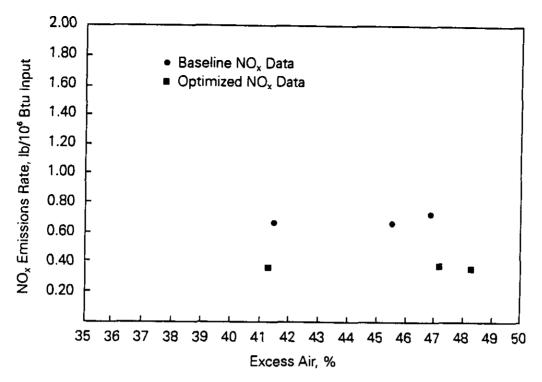


Fig. 19 LNCB™ NO, emissions data for low (54%) load, 4 mills in service.

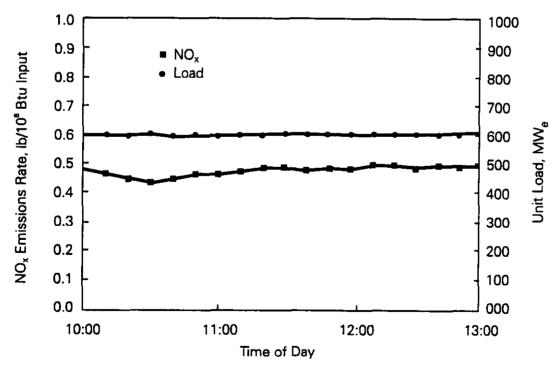


Fig. 20 LNCB™ one-day test results, March 1, 1993.



Technical Paper

GAS REBURNING AND LOW NO_x BURNERS ON A WALL-FIRED BOILER

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ABSTRACT

Parametric tests were conducted for a Gas Reburning-Low NO_X Burner system on a 172 MWe (gross) wall-fired boiler. At 150 MWe net load, the initial low NO_X burner design reduced NO_X emissions from 0.73 lb/10⁶ Btu (314 mg/MJ) to 0.50 lb/10⁶ Btu (215 mg/MJ), a 31 percent overall reduction. At the same net load, with Gas Reburning-Low NO_X Burner operation using 20 percent of total heat input provided by natural gas, NO_X emissions were reduced further to 0.20 lb/10⁶ Btu (86 mg/MJ), a 72 percent overall reduction. These short-term NO_X emissions remained fairly constant when gas heat input ranged from 16 to 23 percent. NO_X emissions decreased linearly with decreasing excess air level at the boiler exit. At baseline or pre-LNB, GR conditions, CO was less than 200 ppm. Baseline carbon loss was less than 6 percent carbon in the ash. The Gas Reburning-Low NO_X Burners operation and the Low NO_X Burners operation produced CO and carbon in ash in these ranges. The heat rate was increased by about 1 percent in the Gas Reburning-Low NO_X Burner operation. Long-term demonstration testing based on automatic, load-following operation started in April 1993 and initial long-term NO_X results agreed with the parametric test results at the same excess air levels.

INTRODUCTION

A Gas Reburning system combined with low NO_X burners was installed and is being evaluated on a 172 MWe (gross) wall-fired utility boiler. The objective of this project ^[1] is to demonstrate that the combination of Gas Reburning (GR) and low NO_X burners (LNB) will achieve 70 to 75 percent NO_X reduction. This \$16.2 million project is a Clean Coal Technology III program sponsored by the U.S. Department of Energy, Gas Research Institute, Public Service Company of Colorado, Colorado Interstate Gas, Electric Power Research Institute, and Energy and Environmental Research Corporation (EER). The GR system including an overfire air system was designed and installed by EER. The LNB system was designed and installed by Foster Wheeler. The parametric testing of the GR-LNB system has been completed. Long-term demonstration testing of the system is currently in progress to determine its impacts on the boiler and boiler operation.

With GR about 80 to 85 percent of the primary fuel is fired in the primary burner zone. The balance of the heat input is provided downstream by natural gas. The gas is injected into the furnace above the primary coal burner zone to produce a slightly fuel-rich zone where NO_X produced by the coal combustion is "reburned" and reduced to atmospheric nitrogen. Combustion is completed by the addition of overfire air (OFA). GR also reduces SO₂, particulates, and CO₂ (a greenhouse gas) by about 20, 20, and 8 percent, respectively, as a result of 20 percent substitution of the heat input by natural gas. This is because natural gas does not contain sulfur or ash and has a higher hydrogen/carbon ratio than coal. The level of NO_X reduction achievable with GR using 15 to 20 percent natural gas is approximately 50 to 60 percent. The NO_X reduction goal for the combination of GR and LNB technologies is 70 to 75 percent.

The host boiler for the project is Cherokee Station Unit 3 at Denver, Colorado. It is owned and operated by Public Service Company of Colorado (PSCO). The unit fires Colorado bituminous coals. The LNB system consists of 16 Foster Wheeler Internal Fuel Staging burners.

This paper describes the boiler, GR-LNB technology, parametric test results, and initial long-term test results.

BOILER DESCRIPTION

Cherokee Station Unit 3 is a 172 MWe (gross) front wall-fired electric facility (Figure 1) located in Adams County, Colorado. The boiler is a balanced-draft pulverized-coal unit supplied by Babcock & Wilcox. As the demand load for the station rises, load on each of four units increases proportionally. Individual units are loaded incrementally based upon current heat rates. The capacity factor and swing load conditions allow evaluation of GR-LNB performance over a wide range of boiler operating conditions with minimal impact on normal plant operations.

Low-sulfur coal (typically 0.4 percent sulfur) is fed to four Riley Stoker No. 556 duplex drum type coal breaker and pulverizing mills, each having a maximum capacity of 37,000 lb/hr (16,800 kg/hr). Coal fed to the mills is pulverized so that at least 70 percent will pass through 74 micron

openings (a 200 mesh U.S.S. sieve) and at least 98.5 percent will pass through 297 micron openings (a 50 mesh U.S.S. sieve).

The pulverized coal is transported to a 4x4 array of Foster Wheeler Internal Fuel Staging low NO_x burners, located on the front wall of the boiler. The radiant zone is 24 ft (7.3m) deep and 42 ft (12.8m) wide and has a full division wall. At the original full load, the design heat input is 1.65 x 10^9 Btu/hr (1,740 GJ/hr). Natural gas was available at the plant prior to the project. Baghouses are used to control particulate emissions to less than 0.1 lb/ 10^6 Btu (43 mg/MJ).

GR-LNB TECHNOLOGY

The combined GR-LNB system^[2] is shown schematically in Figure 2. Several recent references on gas reburning are available ^{[3]-[10]}.

Low NO, Burners

Sixteen Foster Wheeler Internal Fuel Staging low NO_x burners replaced the Babcock & Wilcox circular-type PL burners. The LNBs employ dual combustion air registers which allow for control of air distribution at the burner, providing independent control of the ignition zone and flame shaping. These are designed to achieve a goal of 45 percent reduction of NO_x at 150 MWe (net) relative to the nominal baseline emission of 0.73 lb/10⁶ Btu (314 mg/MJ) at 20 percent excess air or 3.5 percent O_2 on a dry basis.

Gas Reburning System

Natural gas, the reburning fuel, is injected together with recirculated flue gas (FGR) through sixteen 5.5 inch (14.0 cm) diameter front and rear wall nozzles - eight located on each wall. Approximately 3.4 percent of the flue gas is injected through the gas reburning nozzles to improve mixing of natural gas and dispersion within the furnace. This configuration provides for adequate wall to wall and lateral dispersion. The nozzle exit velocity varies linearly with boiler load, ranging from about 90 ft/sec (27.4 m/s) at 50 percent load to 180 ft/sec (54.9 m/s)

at 100 percent load. At full load, the required velocity head for the composite nozzles is 4 inches (10.2 cm) of water column. The range of design flow rates of natural gas is 10 to 25 percent of the total heat input. The natural gas is transported by means of the flue gas constituting 3 to 4 percent of the total flue gas and injected through ports above the upper row of burners. The injection velocity is kept low to minimize furnace flow disruption.

Overfire Air

Overfire air is injected into the furnace through six 20.5 inch (52.1 cm) diameter injectors located on the front wall of the furnace. The injectors are tilted downward 10 degrees to improve overfire air dispersion and to increase residence time. The amount of air added at this point is to complete burning of residual natural gas and bring excess air levels to non-GR values.

Operation

To begin operation of the GR system, the operator first starts the overfire air booster fan. Then the overfire air flow is increased until the desired primary burner zone stoichiometry (Figure 2) is achieved. After selecting a reburning zone stoichiometry, natural gas flow is manually initiated and then switched to automatic control of gas reburning. To shut down the GR system, the operator reverses these steps. While the system is being shut down, cooling air is fed through all GR nozzles.

Safety System

The GR system functions independently of the boiler in that a GR system trip will not trip the boiler. Interlocks are designed to start the GR equipment in an orderly fashion and prevent the operator from allowing the unit's safety to become compromised either through erroneous operation or due to equipment failure. All major commands issued by the control system are verified by a feedback signal. Trip signals are continuously monitored by the control system and will prevent startup or shutdown equipment already in operation.

The GR process does not produce a luminous flame capable of being sensed by conventional flame scanners. To insure that natural gas is not injected into a cold furnace, 8 flame scanners monitor the presence of main fuel flames in the boiler. Loss of signal from 4 scanners automatically shuts down the GR system.

STOICHIOMETRIC RATIOS

The GR process (Figure 2) can be best described by considering three combustion zones in series:

- Primary burner zone: approximately 80-85 percent of the heat is released by coal in this zone under low excess air conditions, achieving a small reduction of NO_x.
- Reburning zone: the reburning fuel, in this case natural gas (normally 15 to 20 percent of the total heat input), is injected downstream of the primary burner zone in the upper furnace to create a slightly fuel-rich zone where NO_X is reduced to elemental N₂.
- Burnout zone: in the third and final zone, additional combustion air (overfire air or OFA) is added to burn any remaining fuel fragments and complete the combustion process.

Each of the three zones in Figure 2 has its unique stoichiometric ratio (SR). The three SR values can be calculated from the following equations:

Primary Burner Zone
$$SR_1 = \frac{TA - OFA}{CSA}$$
 (1)

Reburning Zone
$$SR_2 = \frac{TA - OFA}{CSA + GSA}$$
 (2)

Burnout Zone
$$SR_3 = \frac{TA}{CSA + GSA}$$
 (3)

The symbols used in these equations are defined as follows:

TA = Total air, scfm or Nm³/s

OFA = Overfire air, scfm or Nm^3/s

CSA = Coal stoichiometric air, scfm or Nm³/s

GSA = Natural gas stoichiometric air, scfm or Nm³/s

Since there are seven variables in three equations, only four variables are independent variables.

PARAMETRIC TEST RESULTS

The parametric tests were conducted by changing the process variables, such as zone stoichiometries, percent gas input, percent overfire air, FGR, load, etc. The effects of these variables on NO_X reduction, SO₂ reduction, CO emissions, carbon in ash, and heat rate were studied. At full load, the boiler is normally operated with four coal pulverizing mills. Each mill supplies coal to one row of four burners. In the gas reburning operation using natural gas as a reburning fuel at 20 percent of the total heat input, the boiler can be operated with three mills even at full load.

Typical Operation Profile

A typical operation profile is shown in Figure 3. At a constant load (150 MWe) and a constant O_2 level at the boiler exit (not shown in diagram), both NO_x and SO_2 emissions decrease when natural gas is introduced in the GR operation. If natural gas supply is discontinued, NO_x and SO_2 emissions increase, as expected. A similar trend is exhibited by NO_x and SO_2 emissions at 120 MWe. When the load is decreased from 150 to 120 MWe, NO_x emission decreases but SO_2 emission in $1b/10^6$ Btu (or mg/MJ) remains unchanged since the latter is dependent only on the sulfur content of coal.

Effect of Stoichiometry

Over several months, extensive parametric tests of GR have been completed at Cherokee. Figure 4 shows the results as a function of zone stoichiometry. For the baseline and LNB tests, which involve a single combustion zone, the stoichiometry is the overall stoichiometry. For GR-LNB,

the stoichiometry refers to the reburning zone. Table 1 shows the NO_X results of the parametric tests. For the baseline and LNBs, the table presents data for 20 percent excess air. For GR-LNB, the table presents data for the minimum NO_X level, at a reburning zone stoichiometry of 88 percent of theoretical air. At this point, the gas heat input was 20 percent. The minimum NO_X emission with GR-LNB measured to date was 0.20 lb/10⁶ Btu (86 mg/MJ). This corresponds to a NO_X reduction of 72 percent from baseline levels and 60 percent reduction from using only the low NO_X burners.

As listed in Table 1, present NO_x reduction with LNB operation is 31 percent. Foster Wheeler plans to make burner revisions during the planned January, 1994 boiler outage to achieve the goal of 45 percent NO_x reduction.

TABLE 1

NO_x Data from Cherokee Unit 3: Parametric Tests

Firing Configuration	NO Em lb/10 ⁶⁸ Btu	issions <u>(mg/MJ)</u>	NO _x Redu Baseline	nction (%) Relative to: Low NO _x Burners
Baseline	0.73	(314)	0	NA
Low NO _x Burners (initial design)	0.50	(215)	31	0
Gas Reburning and Low NO _x Burners	s 0.20	(86)	72	60

Effect of Excess Air

Figure 5 shows NO_x emissions vs. percent O_2 dry at the boiler exit. A linear relationship was obtained between NO_x and O_2 for baseline, LNB, and GR-LNB.

Effect of Gas Heat Input

In general, the NO_x emission is reduced with increasing gas heat input, as shown in Figure 6. At gas heat inputs greater than 10 percent, the NO_x emission is reduced marginally with

increasing gas heat input. It looks like that 10 percent gas heat input is optimal for NO_X reduction per unit gas heat input.

Natural gas also reduces SO_2 emissions in proportion to the gas input. At Cherokee Station, low sulfur coal is used and the typical SO_2 emissions are 0.65 lb/10⁶ Btu (280 mg/MJ). With a gas heat input of 20 percent, SO_2 emissions are decreased by 20 percent to 0.52 lb/10⁶ Btu (224 mg/MJ), as expected from fuel substitution by natural gas essentially free from sulfur.

The CO_2 emission is also reduced as a result of using natural gas because natural gas has a lower carbon/hydrogen ratio than coal. At Cherokee, CO_2 emissions from typical coal and natural gas combustion are 210 lb/ 10^6 Btu (90.3 g/MJ) and 120 lb/ 10^6 Btu (51.6 g/MJ), respectively. At a gas input of 20 percent, the CO_2 emission is reduced by 8 percent.

Effect of Load

The effect of load on NO_X is shown in Figure 7. For baseline, LNB, and GR-LNB, the NO_X emission increases with increasing load. The increase in NO_X with increasing load is more moderate with GR-LNB than that with baseline or LNB as indicated by the slopes of the curves.

CO Emissions and Carbon Loss

Baseline CO is less than 200 ppm. Baseline carbon loss is less than 6 percent carbon in ash. Both baseline CO and baseline carbon loss decrease with increasing excess air level at the boiler exit. The CO and carbon are converted to CO₂ more readily at a higher excess air level. Both LNB and GR-LNB produced CO and carbon in ash in these ranges under similar or lower excess air conditions.

Heat Rate

The factors that affect the heat rate are

- Carbon loss
- Dry gas loss (related to excess air and boiler exit temperature)
- Latent heat loss (related to H₂O in the combustion products)
- Steam temperature (affecting turbine cycle efficiency)
- Auxiliary power

The carbon loss remains unchanged with GR operation. The dry gas loss is essentially unaffected because an increase of about 10°F (6°C) in boiler exit temperature is canceled out by a reduction in excess air. The latent heat loss reduces the boiler efficiency by about 1 percent when using 20 percent gas heat input. The steam temperature can be maintained via attemperation. The slight increase in auxiliary power use is offset by the reduced mill power. Overall, the heat rate increased about 1 percent.

Data Prediction

Based on the parametric test results, the BrainMaker (a neural network that can "learn" from experience and make predictions) predicted NO_x levels which agreed with measured NO_x levels during the long-term testing, as shown in Figure 8. The four major independent process variables used in the BrainMaker for NO_x prediction are load, CEMS O_2 , gas heat input, and reburning zone stoichiometry (SR₂) for LNB, LNB-OFA, and GR-LNB operations.

LONG-TERM TESTING

Long-term testing started in the last week of April, 1993 and will last for one year. The objective of the testing is to obtain operating data over an extended period when the unit is under routine normal commercial service, determine the effect of GR-LNB operation on the unit and obtain the incremental maintenance and operating costs with GR.

Operating Data

The nominal long-term testing conditions specified are a primary burner zone stoichiometry of 1.08, a burnout zone stoichiometry of 1.18, a gas heat input of 18 percent, and the FGR flow rate of 10,000 scfm (4.7 Nm³/s), based on the parametric test results. However, the gas heat input will be lowered to 10 percent or so during some periods of the long-term testing.

The initial long-term test results obtained in the first three months agreed with the parametric test results. The long-term data fall on the same curves of NO_x vs. CEMS O_2 (O_2 dry measured with a Continuous Emissions Monitoring System at the boiler exit) and NO_x vs. load in Figures 5 and 7.

Average NO_x reductions (based on the pre-LNB baseline NO_x level of 0.73 lb/10⁶ Btu or 314 mg/MJ) and CEMS O₂ levels in various tests are plotted against test dates in Figure 9. It is seen that the NO_x reduction curve is essentially a mirror image of the CEMS O₂ curve. This means that a higher NO_x reduction is achieved at a lower CEMS O₂, as also shown in Figure 5. As usual, it is necessary to maintain CEMS O₂ at or slightly less than 3% O₂ on a dry basis to achieve the greatest NO_x reduction. However, if the CEMS O₂ level is too low, CO will increase exponentially. The average long-term NO_x reduction achieved to date is 64% (ranging from 54% to 72%), reflecting the variability in CEMS O₂. This variability, in turn, is a result of boiler operation where O₂ is controlled manually.

Costs Data

Maintenance and operating costs associated with the GR operation will be obtained over the testing period. While equipment costs can be determined to a fairly precise level, only operation of the system can establish operating and maintenance costs. Accordingly, a system has been set up that will gather pertinent cost data over the 12-month testing period.

The GR system was designed on the basis of using 18 to 20 percent natural gas input. Parametric testing has shown substantial NO_x reduction at 10 percent natural gas input. As the

size of GR equipment will be smaller at 10 percent gas heat input, both capital and operating costs will be lower. The unit will be operated with 10 percent gas input during part of long-term testing.

A preliminary cost estimate indicates that the natural gas cost (in \$/ton NO_x removed) in the GR process can be decreased by approximately 30 percent in lowering the gas heat input from 18 to

10 percent. This natural gas cost is estimated at a NO_x reduction ratio of 1:0.8 (based on LNB NO_x) with 18 and 10 percent gas heat inputs.

SUMMARY

- 1. Parametric tests were conducted for a Gas Reburning-Low NO_X Burner system on a 172 MWe (gross) wall-fired boiler. At 150 MWe net load and 20 percent excess air, NO_X emissions from the wall-fired boiler were reduced from 0.73 lb/10⁶ Btu (314 mg/MJ) to 0.50 lb/10⁶ Btu (215 mg/MJ) by low NO_X burners (a 31% reduction). NO_X emissions were reduced to 0.20 lb/10⁶ Btu (86 mg/MJ) (a 72% overall reduction) by Gas Reburning with 20 percent gas input combined with the low NO_X burners.
- 2. The NO_X level remained fairly constant when gas input was increased from 16 to 23 percent.
- 3. NO_X decreased linearly with decreasing excess air level at the boiler exit. With this boiler/burner/fuel combination, the boiler exit excess air must be maintained at 3% O₂ dry or slightly lower to maximize NO_X reduction while maintaining a reasonably low CO level.
- Baseline CO was less than 200 ppm. Baseline carbon loss was less than 6 percent carbon in ash. The Gas Reburning-Low NO_X Burners operation produced CO and carbon in ash in these ranges.

- 5. The heat rate was increased by about 1 percent with the gas reburning-low NO_X burner operation.
- Long-term demonstration test results for NO_X levels obtained to date at constant or variable loads are in good agreement with parametric test results at the comparable excess air levels.
- 7. The BrainMaker program can predict NO_X levels within experimental errors after having been trained with data points.
- 8. The average NO_x reduction obtained to date in long-term testing is 64%, compared to 70% as the goal. The lower NO_x reduction value is a result of the variability in CEMS O_2 largely due to the manual operation of the boiler.

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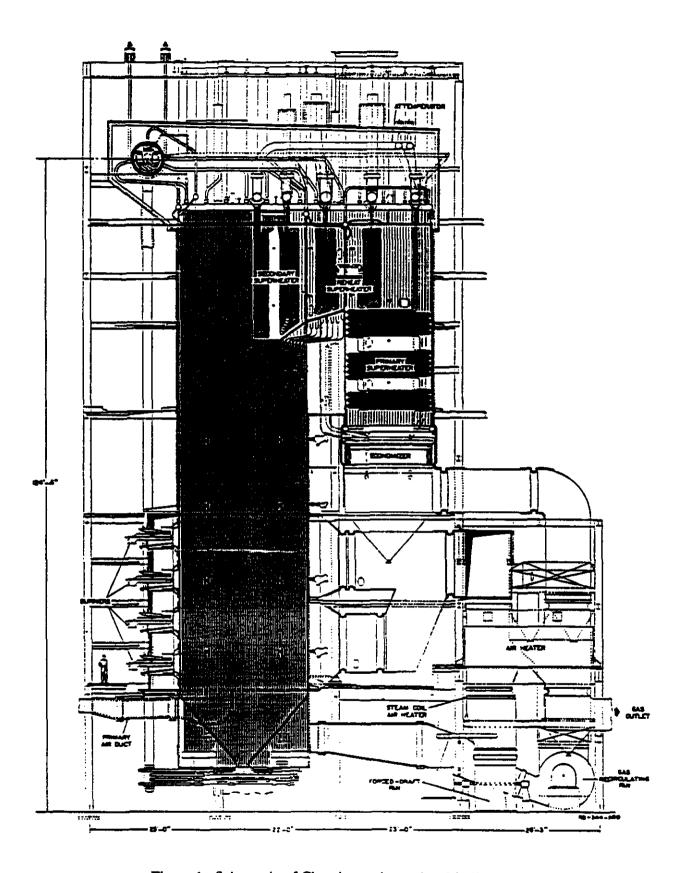


Figure 1. Schematic of Cherokee unit number 3 boiler.

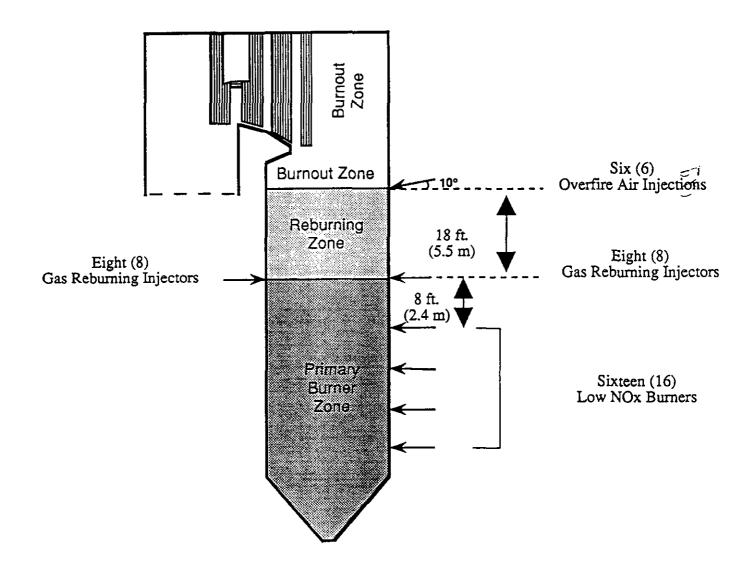


Figure 2. GR-LNB system schematic at Cherokee.

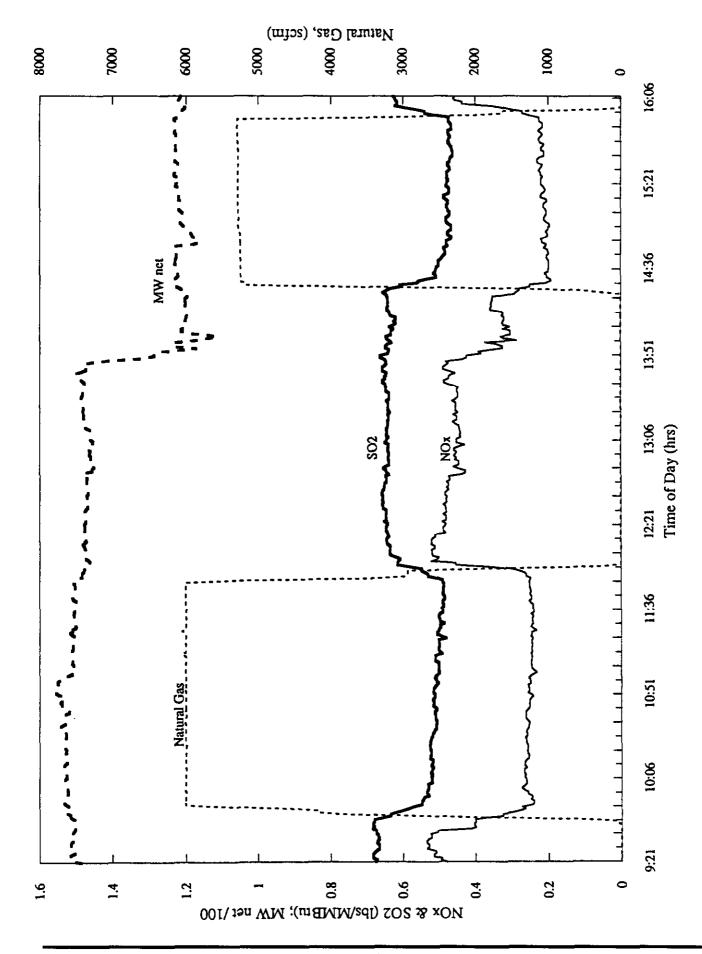


Figure 3. Typical operation profile (1 lb/MMBtu = 430 mg/MJ, MMBtu and MJ based on gross or higher heating values, 1 scfm = $4.72 \times 10^4 \text{ Nm}^3/\text{s}$).

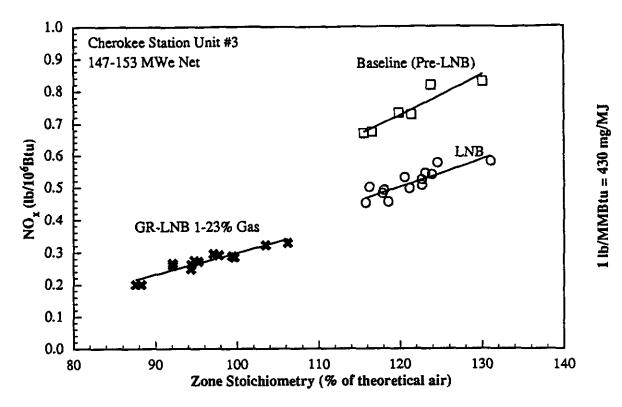


Figure 4. Gas reburning data from Cherokee.

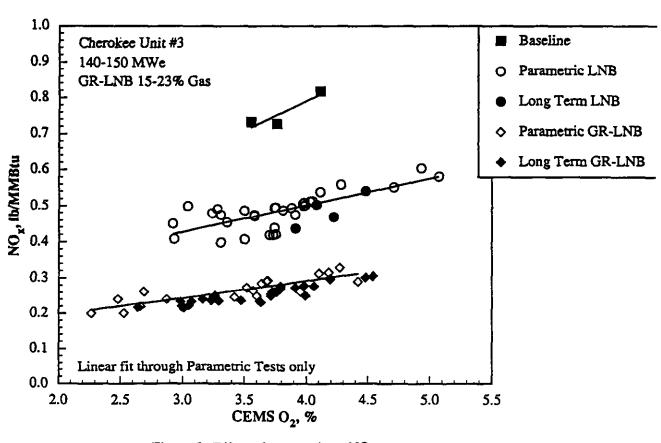


Figure 5. Effect of excess air on NOx.

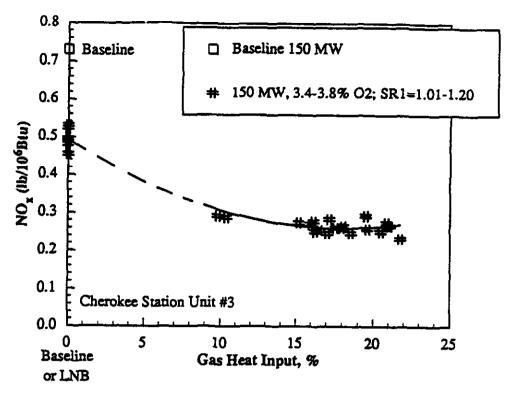


Figure 6. Effect of gas inupt on NO_x.

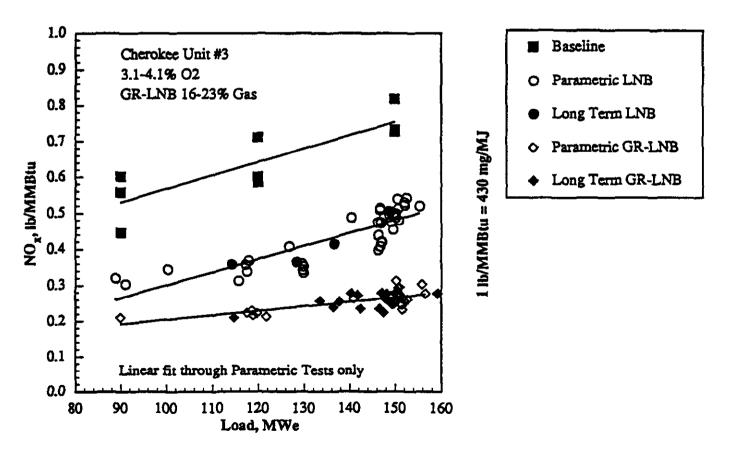


Figure 7. Effect of load on NO_x.

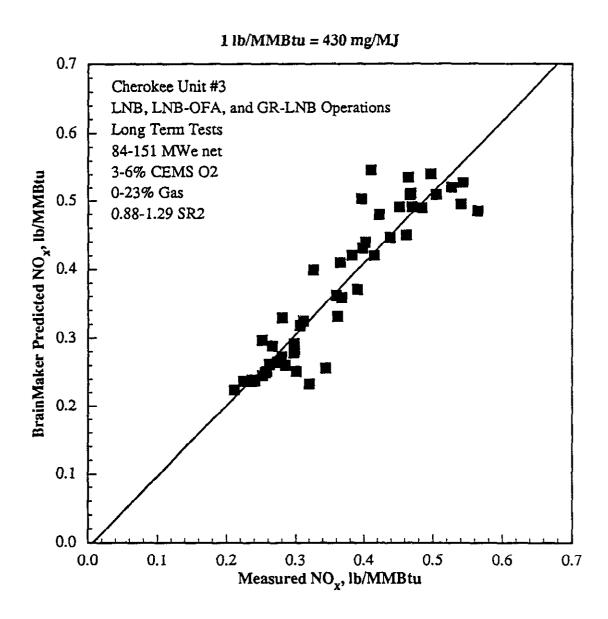


Figure 8. Predicted NO_x vs. measured NO_x .

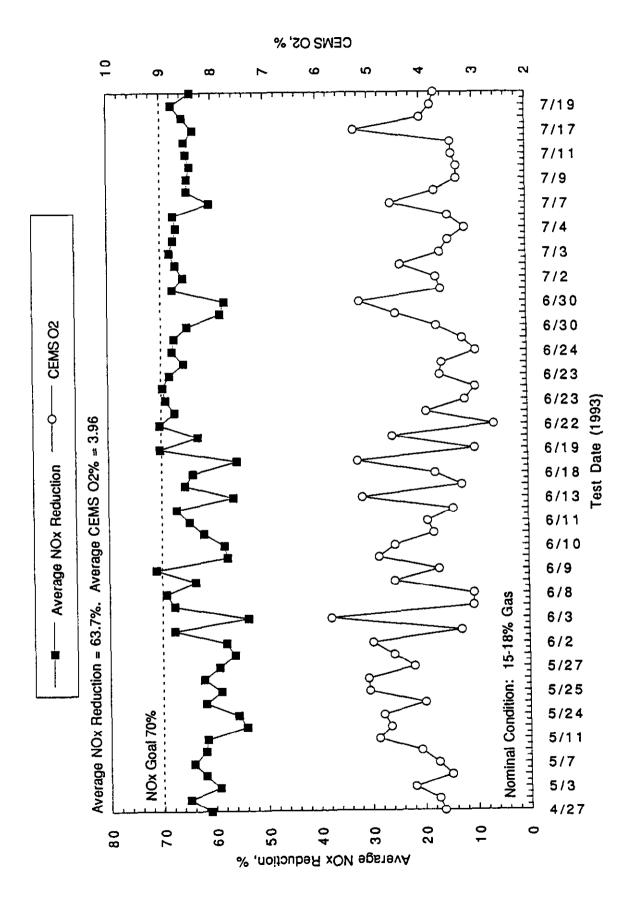


Figure 9. Long term GR-LNB test performance at Cherokee.



DEMONSTRATION OF SELECTIVE CATALYTIC REDUCTION TECHNOLOGY FOR THE CONTROL OF NITROGEN OXIDE EMISSIONS FROM HIGH-SULFUR, COAL-FIRED BOILERS

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ABSTRACT

This paper describes the status of the Innovative Clean Coal Technology project to demonstrate SCR technology for reduction of NOx emissions from flue gas of utility boilers burning U.S. high-sulfur coal. The funding participants are the U.S. Department of Energy (DOE), Southern Company Services, Inc. (SCS), on behalf of the entire Southern Company, Electric Power Research Institute (EPRI), and Ontario Hydro. SCS is the participant responsible for managing all aspects of the project. The project is being conducted on Gulf Power Company's Plant Crist Unit 5 (75-MW nominal capacity), located near Pensacola, Florida, on U.S. coals that have a sulfur content near 3.0%. The SCR facility treats a 17,400 scfm slip-stream of flue gas and consists of three 2.5-MW (5000 scfm) and six 0.2-MW (400 scfm) SCR reactors. The reactors operate in parallel with commercially available SCR catalysts obtained from vendors throughout the world. The design engineering and construction have been completed, and the start-up/shakedown was completed in June 1993. Long-term performance testing began in July 1993 and will be conducted for two years. Test facility description and test plans, as well as start-up issues and preliminary commissioning test results are reported in this paper.

DEMONSTRATION OF SELECTIVE CATALYTIC REDUCTION TECHNOLOGY FOR THE CONTROL OF NITROGEN OXIDE EMISSIONS FROM HIGH-SULFUR, COAL-FIRED BOILERS

INTRODUCTION

The need within the utility industry for detailed information on selective catalytic reduction (SCR) technology has never been greater. The 1990 Clean Air Act Amendments (CAAA) create two new nitrogen oxide (NO_X) control requirements on fossil fuel-fired utility boilers. First, Title IV of the CAAA regarding acid rain requires that emission limits be placed on all coal-fired utility boilers in two phases, one beginning in 1995 and the other in the year 2000. SCR, in which ammonia is added to the flue gas to reduce NO_X to nitrogen over a catalyst, is not as prominently mentioned as low NO_X burner technology for meeting the Title IV provisions. However, the final EPA emission limitations for each of the two phases remain to be established, and SCR is still very much under consideration in utilities' compliance strategies. Second, Title I of the CAAA addresses attainment of the ambient air quality standards. Regarding ozone, Title I calls for certain areas presently not in attainment to consider NO_X controls to achieve attainment. As a result, renewed focus has been placed on NO_X controls, including advanced NO_X control technologies such as SCR, which may be required to meet compliance requirements for ozone non-attainment areas.

SCR technology involves the injection of ammonia into flue gas and then passing the gases through one or more catalyst layers where NO_X and ammonia react to form nitrogen and water vapor. A simplified, typical SCR process installation for a utility boiler is depicted in Figure 1. Hot flue gas leaving the economizer section of the boiler is ducted to the SCR reactor. Prior to entering the reactor, ammonia (NH₃) is injected into the flue gas at a sufficient distance upstream of the SCR reactor to provide for complete mixing of the NH₃ and flue gas. The quantity of NH₃ is adjusted to achieve the desired NO_X removal efficiency. The reactions between NH₃ and NO_X occur as the flue gas passes through the catalytic layers of the SCR reactor. Ductwork is installed to bypass some flue gas around the economizer during periods when the boiler is operating at reduced load. This is done, especially on retrofits, to maintain the temperature of the flue gas entering the catalytic reactor at the proper reaction temperature of about 700°F.

SCR technology is in commercial use in Japan and Western Europe on gas-, oil-, and low-sulfur, coal-fired power plants. There are now over 36,000 MW of fossil-fuel-fired SCR capacity in

Japan, including 6,200 MW on coal. There are over 33,000 MW of fossil-fuel-fired SCR capacity in Western Europe, including 30,500 MW of coal-fired capacity.¹

SCR DEMONSTRATION GOALS

Although SCR is widely practiced in Japan and Western Europe, numerous technical uncertainties are associated with applying SCR to U.S. coals. These uncertainties include:

- (1) potential catalyst deactivation due to poisoning by trace metal species present in U.S. coals but not present, or present at much lower concentrations, in fuels from other countries;
- (2) performance of the technology and effects on the balance-of-plant equipment in the presence of high amounts of SO₂ and SO₃ (e.g., plugging of downstream equipment with ammonia-sulfur compounds); and
- (3) performance of a wide variety of SCR catalyst compositions, geometries and manufacturing methods at typical high-sulfur coal-fired utility operating conditions.

These uncertainties are being explored by constructing and operating a series of small-scale SCR reactors and simultaneously exposing different SCR catalysts to flue gas derived from the combustion of high-sulfur U.S. coal. The first uncertainty will be handled by evaluating SCR catalyst performance for two years under realistic operating conditions found in U.S. pulverized-coal-fired utility boilers. Deactivation rates for the catalysts exposed to flue gas of high-sulfur U.S. coal will be documented to determine catalyst life and associated process economics. The second uncertainty will be explored by performing parametric tests, during which SCR operating conditions will be adjusted above and below design values to observe deNO_X performance and ammonia slip. The performance of air preheaters installed downstream of the larger SCR reactors will be observed to evaluate the effects of SCR operating conditions upon heat transfer and boiler efficiency. The third uncertainty is being addressed by using honeycomb- and plate-type SCR catalysts of various commercial compositions from the U.S., Japan, and Europe. Tests with these catalysts will expand knowledge of the performance of SCR catalysts under U.S. utility operating conditions with high-sulfur coal.

The intent of this project is to demonstrate commercial catalyst performance and to determine optimum operating conditions and catalyst life for the SCR process. This project will also demonstrate the technical and economic viability of SCR while reducing NO_X emissions by at least 80%.

SCR DEMONSTRATION FACILITY DESCRIPTION

The SCR demonstration facility is located at Gulf Power Company's Plant Crist in Pensacola, Florida. The facility will treat a flue gas slip-stream from Unit 5, a commercially operating 75-MW unit, firing U.S. coals with a sulfur content near 3.0%. Unit 5 is a tangentially-fired, dry bottom boiler with hot- and cold-side electrostatic precipitators (ESPs) for particulate control. The SCR test facility consists of nine reactors operating in parallel for side-by-side comparisons of commercially available SCR catalysts obtained from vendors throughout the world. With all reactors in operation, the amount of combustion flue gas that can be treated is 17,400 scfm or 12% of Unit 5's capacity (about 8.7 MWe).

The process flow diagram for the SCR test facility is shown in Figure 2. There are three large SCR reactors (2.5 MW, 5000 scfm) and six smaller SCR reactors (0.2 MW, 400 scfm). Eight of the nine reactors will operate with flue gas containing full particulate loading (high dust) extracted from the inlet duct of the hot-side ESP, while one small reactor will use flue gas fed from the ESP outlet (low dust).

Each reactor train has electric duct heaters to control the temperature of the flue gas entering the reactor and a venturi flow meter to measure the flue gas flow. An economizer bypass line to the SCR test facility maintains a minimum temperature of 620°F for flue gas supplied to the test facility. Anhydrous ammonia is independently metered to a stream of dilution air that injects the ammonia via nozzles into the flue gas stream prior to each SCR reactor. The flue gas and ammonia pass through the SCR reactors, which have the capacity to contain up to four catalyst layers.

For the large reactor trains, the flue gas exits the reactor and enters a pilot-scale air preheater (APH). The APHs are incorporated in the project to evaluate the effects of SCR reaction chemistry on APH deposit formation and the effects of the deposits on APH performance and operations. All reactor trains, except the low-dust train, have a cyclone downstream of the SCR reactor to protect the induced draft (ID) fan from particulates. The exhaust for all the SCR reactors is combined into a single manifold and reinjected into the host boiler's flue gas stream ahead of the cold-side ESP. The preheated air from the APH on the large reactors is also combined into a single manifold and returned to the host boiler draft system at the air outlet of the existing APH. All of the particulates that are removed from the flue gas with the cyclones are combined and sent to an ash disposal area.

CATALYST TESTING PLANS

Seven catalyst suppliers are participating in this project, providing nine different catalysts. The two suppliers from Europe and two from Japan provide one catalyst each. The three U.S. firms are supplying five of the catalysts. The catalysts being evaluated represent the wide variety of SCR catalysts being offered commercially and possess different chemical compositions and physical shapes. Of these nine catalysts, six have a honeycomb geometry while the remaining three are plate-type catalysts. The suppliers, corresponding reactor size, and catalyst configuration are listed in Table 1.

After start-up, the baseline performance of each catalyst will be determined at design conditions which will be maintained for the two year test period. Once baseline performance has been established, each reactor will be sequenced through a test matrix (parametric tests) that varies the following variables around the SCR process design point: ammonia-to-NO_X ratio, temperature, and space velocity. Space velocity is the ratio of flue gas volumetric flow rate to catalyst volume. With a fixed catalyst volume, variations in flue gas flow rates will alter the space velocity around the design point.

DeNO_X efficiency, pressure drop, SO₂ oxidation, and ammonia slip will be determined at each parametric test condition. Once a parametric test matrix has been completed, each reactor will be returned to baseline design conditions. This allows for steady-state operation over a three month period between parametric tests for aging of the catalyst. The parametric test matrix will be repeated every three months for each reactor train. Only one reactor train will be undergoing parametric testing at any one time. The remaining reactors will be either in steady-state operation or off-line. The APH is bypassed during parametric testing so that long-term deposit formation is not affected.

The operating parameter ranges to be examined during the parametric tests and the long-term design condition (baseline) are as follows:

	<u>Minimum</u>	Baseline	<u>Maximum</u>
Temperature, °F	620	700	750
NH ₃ /NO _x molar ratio	0.6	0.8	1.0
Space velocity,			
 % of design flow 	60	100	150
 Flow rate, scfm 			
-large reactor	3000	5000	7500
-small reactor	240	400	600

PROJECT SCHEDULE AND STATUS

The demonstration project is organized into three phases. Phase I consisted of permitting, preparing the Environmental Monitoring Plan and preliminary engineering. Phase II included detailed design engineering, construction, and start-up/shakedown. Detailed design engineering began in early 1991 and concluded in December, 1992. Construction began at the end of March 1992 and was completed by the end of February 1993. Start-up/shakedown concluded in June 1993. Baseline commissioning tests without catalysts were conducted through June. The loading of all catalysts was completed at the end of June.

The operations phase for process evaluation, Phase III, commenced in July 1993. The process evaluation will last for two years and will be followed by preparation of a final report, which will include process economic projections. The major milestones on the schedule are shown in Table 2.

START-UP ISSUES

As may be normally expected, there have been several problems encountered upon start-up, some of which are not associated with the SCR process per se. The major experiences are highlighted below:

Dilution/Extraction Gas Sampling/Monitoring System

The SCR test facility uses a dilution/extraction sampling system for measurement of NO_X , SO_2 , CO_2 , and CO in the flue gas. This sampling method uses dry air as a dilution medium, with typical air/sample dilution ratios ranging from 100 to 250, to minimize the difficulties associated with the transport and measurement of these gases as compared to other available methods. Problems experienced with this system include accurate measurement of NO_X when ammonia is injected, coordination of the shared analyzers, and communications with the test facility data collection system.

Although the inlet NOx readings are not affected, there have been problems with NO_X measurements at intermediate reactor levels in the presence of ammonia. Apparently catalytic reactions are proceeding in the sampling system, resulting in reduced NO_X values. There has been a series of traps and filters installed in sample lines to capture the ash, water vapor and acid

condensate in order to improve the accuracy of the analyzer system. Work is underway to investigate the use of alternate materials of construction for the sampling probes.

For the nine reactors, there are three NO_X analyzers for the reactor outlet measurements. Each of these analyzers operate on a time-shared basis serving three specific reactors. These systems use a complex system of pumps and valves to direct the sample that is continuously extracted to the analyzer. While one of the three reactor sampling points is active, the other two points are expected to hold their previous values. However, erroneous data is being transmitted for the two points which are supposedly inactive.

The gas analyzer system has a dedicated programmable controller that collects the data from all the analyzers and then sends them to the test facility's control and data collection system. Because these are different systems, the communication protocol had to be worked out during start-up. Although many of the communication problems were solved during the start-up of the test facility, there are still some communication failures occurring. All of these problems with the gas sampling/analysis systems are being addressed.

Ammonia Injection Flow Control

The ammonia vapor flow rates for injection into the reactors are being controlled by precision mass flow control valves. These controllers are affected by liquid in the flow stream, pressure variations, trash in the line, and also the orientation of the controller itself. These controllers were calibrated on nitrogen and scaled to read ammonia flow. Although initial results indicated accurate flow control, subsequent measurements have indicated that actual ammonia flow has been 10 to 25 percent higher than the controllers are indicating. Actions taken to correct this situation include installing coalescent filters on the ammonia supply lines to each control valve, reorienting the controllers, replacing the ammonia header pressure regulator, cleaning each controller, and recalibrating and verifying with other instruments.

Sulfate Deposition

There have been problems with plugging in ductwork where continuous flow is not maintained. These areas provide condensation sites which is exacerbated by the high sulfur concentrations in the fuel and the flue gas. While the ammonia injection system was being completed and flue gas was being passed through the system for startup, the installed injectors presented one such low flow area that sulfates diffused into and precipitated out, plugging almost every injection system.

The nozzles and injection header were cleaned and some portions of the feed piping had to be replaced. The air fan for ammonia dilution has since been placed in service and will be used to supply a continuous air flow to act as a purge to prevent recurrence of the plugging. The horizontal sections of the large reactor bypass lines accumulated a large amount of sulfate formation that blocked operation of several dampers. These dampers are being exercised on a weekly basis to prevent the blockage from binding the dampers again.

Low Dust Reactor Fouling

After only a few hours of operation during its first start-up after catalyst loading, the low-dust reactor experienced severe plugging of the first catalyst layer. While the large reactor bypass lines may be used to flush any ash accumulations associated with the main extraction scoop, the low dust reactor ductwork was not provided with any bypass capability. Also, the isolation damper for that line is approximately 100 feet downstream of the scoop allowing a deadleg for sulfate formation when the reactor is off-line. So during start-up an unusually large amount of solid material may have been introduced to the low-dust reactor. The first layer catalyst element has been returned to the catalyst vendor for examination and a study is underway to evaluate solutions to prevent recurrence of this problem.

Bypass Heat Exchangers

The bypass heat exchangers, which were included for use during the parametric testing on the large reactors to minimize effects of high ammonia slip upon the long-term evaluation of the air preheaters, have been easily plugged by ash and sulfate deposits. Cleaning with either air or water has not been a satisfactory solution. Work is underway to develop another means to cool the flue gas while bypassing the air preheaters.

Ash Accumulation

During start-up, especially during low flows, ash build-up was found in several areas of the ductwork including the main scoop area, the electric flue gas heaters, and the bypass heat exchangers. Extra access ports for soot blowing were added to clean these areas.

Reactor and Air Preheater Soot blowing

Steam soot blowers are used in the large reactor trains for both the catalyst baskets as well as the air preheaters. Much effort has been expended to eliminate the condensate from the soot blowing steam supply piping before the soot blowers extend into the reactors. An extra steam isolation valve has been added on each soot blower and a process steam condensate trap is used on each reactor's steam supply header. Warm-up vents have been added to assure the piping is hot enough to prevent condensation. Follow up inspections reveal that the soot blowers are effective in dislodging any ash build-up on either the reactor baskets or on the air preheater baskets.

Reactor Fans

Due to the small flow, high head requirements of the test facility, the reactor fans are custom designed and not "off the shelf" models. Because of the head requirement, the fan wheels are narrow, large diameter with relatively high inertial moments that made bearing selection difficult. On the small reactor fans, the bearings were replaced twice before changing the design to ball bearings.

Because of the possibility of ammonia slip in the flue gas, materials used in fan construction had to be compatible with ammonia. Ammonia will attack any copper-based alloy. The original vane support bushings were pressed carbon and very brittle; several were broken in shipment and more broke during installation. The first replacements fabricated were brass, and they were rejected due to the ammonia attack of copper alloys. The next offering was stainless steel, which galled as soon as it was installed. The latest solution is a silicon alloyed cast iron, which has performed well over the last three months. The vane bearings have been extended off of the fan housing and new seals have also been installed.

TEST RESULTS

The facility test plan is divided into two main sections, 1) start-up and commissioning tests, and 2) long term testing and parametric evaluation. The start-up and commissioning tests were designed to insure the quality of data obtained from the facility. These tests include base-line evaluations as well as measurements insuring comparability between the reactors. The majority of the tests have been completed and data evaluation is currently underway. The following list describes some of the start-up and commissioning tests that were performed during this section of testing.

- (1) Instrument calibration and gas analysis system verification.
- (2) Base-line particulate concentration, size distribution, and metals concentrations from host unit.
- (3) Base-line chemical composition of host unit slip stream.
- (4) Comparative particulate loading to each reactor.
- (5) SO₂ oxidation characteristics of the system.
- (6) Determination of inherent system ammonia oxidation characteristics.
- (7) Verification of ammonia mass flow control.
- (8) Measurement of catalyst SO2 oxidation characteristics.
- (9) Determination of velocity and particulate profiles at reactor exits.

The following tables and discussions describe some of the most important start-up and commissioning test results that are available at this time. All of the data presented here is of a preliminary nature. Several analyses such as particle size distributions and metals analysis are not available at this time due to the long analytical times required for these measurements.

Table 3 shows the base-line flue gas composition measured in the host unit duct at high (84 MW) and low (43 MW) boiler load. This data compares favorably with data taken several years ago during initial site selection.

Particulate loading in the process stream is a critical design consideration in the development of SCR catalysts. Initial particulate measurements showed that the small reactors were receiving a higher particulate loading than the large reactors under all boiler conditions. After reviewing the design of the splitting section of the main flue gas scoop at the point of the small reactor take-off, the splitting section was mechanically improved to give proper isokinetics, which corrected the particulate loading discrepancies between the reactors. Table 4 gives the particulate loading to each of the eight high dust test facility reactors at high and low boiler load. This data was taken using isokinetic particulate sampling performed as a traverse across the cross-section of the reactor exits. This data compared favorably with the base-line particulate data taken from the host unit duct work.

The data in Table 4 show that the particulate loading to each reactor is fairly consistent and that the loading does not vary more than 10% from the average in most cases. Some of the differences in loading are likely due to boiler variations since individual measurements were taken over a very short period of time with the overall tests taking several weeks. More particulate data

will be obtained as the testing program continues. This should allow long term loading characteristics to be established for each reactor.

Tests have also been performed to determine how evenly the particulates are distributed within the individual reactors. These tests were performed at the reactor exits. Preliminary results indicate that the mass loading is evenly distributed in the cross-sections of the reactors. These measurements were made as six point traverses over the cross-section of the large reactors and three point traverses over the cross-section of the small reactors. Velocity distribution measurements across the reactors at the same sampling locations also indicate a very even velocity distribution.

Sulfur trioxide in the flue gas stream is an extremely important consideration for balance of plant equipment in SCR applications. This is primarily due to the side reaction of SO₃ with ammonia. This reaction forms ammonium bisulfate and sulfate which occur at relatively low temperatures downstream of the SCR reactor, e.g., at the air preheater. SCR catalysts have the potential to oxidize SO₂ to SO₃ thereby exacerbating the ammonium bisulfate/sulfate formation problem as well as contributing to acid deposition problems.

To characterize this oxidation, two series of start-up and commissioning test were performed. The first series of tests characterized the inherent SO₂ oxidation within the test facility system. This included oxidation across the test facility flue gas heaters, as well as oxidation across the reactors themselves (without catalyst). These tests were performed on one large reactor and one small reactor. The results are shown in Table 5. The heater inlet SO₃ values compare favorably with the base line values at low load. However, the high load values for SO3 appear to be considerably lower than base line. This may be due to changes in boiler operation between testing periods (several months). The data show that no net increase in SO₃ was taking place across the SCR reactors. In fact, a slight decrease in SO₃ was noted, which was probably due to deposition in cool spots on the reactor between measurement points. Some oxidation was noted across the flue gas heaters, which was expected. The absolute increase in SO3 over the heaters was greatest at low load. This may be due to the higher heat flux required from the heaters at low unit load to maintain temperature to the SCR reactors. However, the percent increase in SO₃ across the heater at both high and low load is roughly equivalent. The second series of SO2 oxidation tests will determine the oxidative characteristics of the SCR catalysts themselves. These tests will be performed as part of the preliminary parametric sequence. This data is not available at this time.

Upon completion of commissioning tests without catalyst, catalyst loading was completed in late June 1993. Long-term testing and parametric evaluations are underway. Immediately after catalyst loading, all reactors were operated briefly to obtain fly ash samples for the Toxicity Characteristics Leaching Procedure (TCLP) analysis. The TCLP results indicated no detectable amounts or change in constituents between baseline ash samples and ash samples from the SCR process outlet.

The first parametric testing is underway. Based upon the results of this first test, a parametric test plan will be finalized for the remainder of the two year operation of this test facility.

SUMMARY

During this ICCT demonstration, performance data will be developed to evaluate SCR capabilities and costs that are applicable to boilers using high-sulfur U.S. coals. The SCR demonstration facility construction has been completed and start-up/shakedown was finished in early June 1993. Long-term performance testing began in July 1993 and will be completed in 1995.

Operation issues which have been successfully addressed include resolving sulfate deposition in the ammonia injection header system, adding extra soot blower ports to clean areas of ash accumulation, improvements on steam soot blowing of large reactors and air preheaters, and resolving several fan operational issues. Problem areas still being addressed include operation of sampling/monitoring systems, low dust reactor fouling and bypass duct exchanger operation.

In general, the start-up and commissioning tests have demonstrated that each of the SCR reactors is operating on the same basis in terms of process gas feed. Distribution measurements on the individual reactors are in good agreement with the original design requirements. The results of these tests validate the test facility and should guarantee the quality of data obtained in long-term operation and parametric testing.

REFERENCES

1. A. L. Baldwin, J. D. Maxwell, <u>U.S. Department Of Energy's and Southern Company Services's August 24 -September 1, 1991, Visit to European SCR Catalyst Suppliers</u>, U.S. DOE, Pittsburgh, PA, 1991, p 41-3.

Table 1. SCR Project Catalyst Suppliers.

Catalyst Vendor	Reactor Size	Catalyst Configuration
Nr Chalada'	T	TT
Nippon Shokubai	Large	Honeycomb
Siemens AG	Large	Plate
W. R. Grace	Large	Honeycomb
W. R. Grace	Small	Honeycomb
Haldor Topsoe	Small	Plate
Hitachi Zosen	Small	Plate
Cormetech	Small	Honeycomb
Engelhard	Small	Honeycomb (high dust)
Engelhard	Small	Honeycomb (low dust)

Table 2. Project Schedule

Detailed Engineering	1/92 - 12/92
Construction	3/92 - 2/93
Start-up/Shakedown	1/93 - 6/93
Process Evaluation	7/93 - 6/95
Disposition/Final Report	7/95 - 10/95

Table 3. Test Facility Inlet Flue Gas Composition

Constituent	ESP Inlet		ESP Outlet	
	84 MW	<u>43 MW</u>	84 MW	<u>43 MW</u>
NO_X	325	401	332	Not Available
SO ₂ (ppm)	2340	1780	2030	1510
SO ₃ (ppm)	32	42	14	20
HCl (ppm)	104	89	115	101
NH ₃ (ppm)	<0.4	<0.4	<0.4	<0.4
Particulate (gr/dscf)	3.76	2.43	0.0018	BDL*

^{*} Below detection limits

Table 4. Particulate Loading to Reactors

Reactor	Ash Loading (84 MW)	Ash Loading (43 MW)	
	(gr/dscf)	(gr/dscf)	
A	3.65	3.08	
В	4.18	3.04	
С	3.96	3.16	
D	2.83	2.70	
E	3.96	3.22	
F	4.01	3.04	
G	3.60	2.71	
H	3.52	2.75	

Table 5. SO2 Oxidation Across Test Facility Without Catalyst

		SO3 (ppm)		
		Heater Inlet	Heater Exit	Reactor Exit
Large Reactor	84 MW	12	15	10
	43 MW	31	40	32
Small Reactor	84 MW	8	11	7
	43 MW	28	35	23

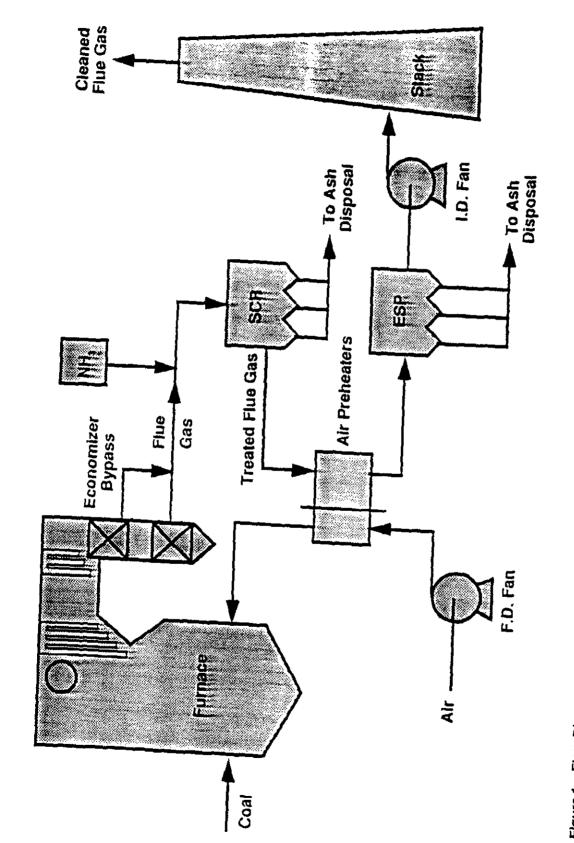


Figure 1. Flow Diagram of a Typical SCR Installation.

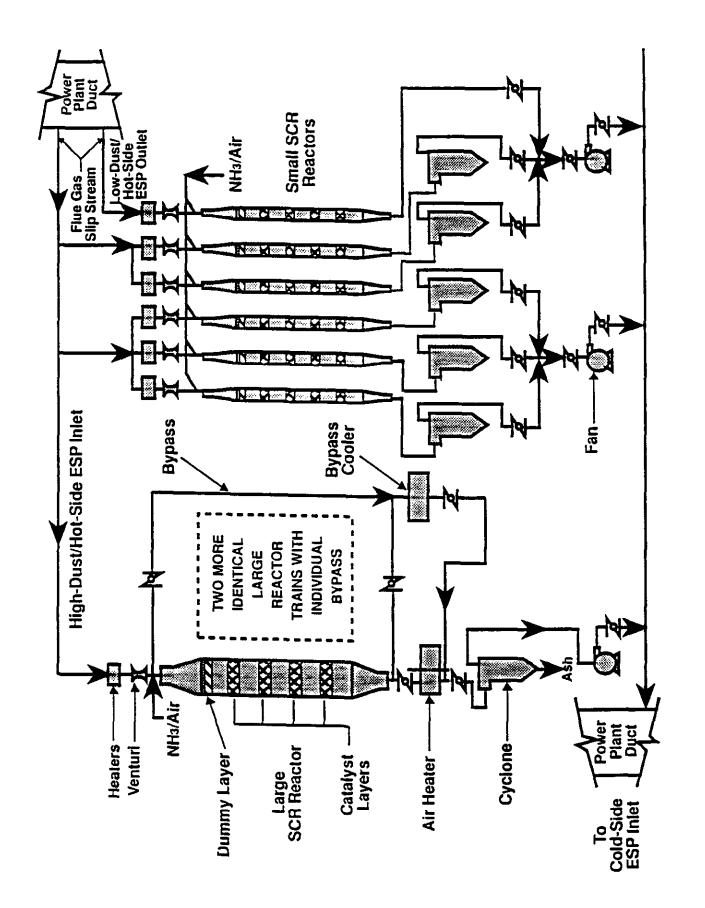


Figure 2. Prototype SCR Demonstration Facility-Process Flow Diagram.

Design Methodology for a Micronized Coal Reburn System Using Modeling

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Presented at

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ABSTRACT

The Tennessee Valley Authority (TVA) has been selected by the Department of Energy's Clean Coal Technology IV program to demonstrate micronized coal reburn technology for control of nitrogen oxide (NO₂) emissions. The demonstration will be done at full scale on a 175 MWe wall-fired steam generator at the Shawnee Fossil Plant. The micronization technology of the Fuller Corporation makes this demonstration feasible, hence, TVA has selected Fuller as the prime contractor for the project and partner in the commercialization of the technology. Radian Corporation has been selected to define the combustion and mixing aspects of the demonstration. Radian Corporation will thus define the design of the reburn injection and OFA system to be installed. This retrofit demonstration is expected to decrease NO, emissions by 50 to 60 percent. Up to 30 percent of the total fuel fired in the furnace will be micronized coal injected in the upper furnace creating a fuel-rich reburn zone. Overfire air will be injected at conditions that will attain good furnace gas mixing above the reburn zone to insure complete combustion. This paper outlines the efforts to be conducted in defining the key parameters associated with injection and mixing of the micronized coal reburning media and the overfire air (OFA). Shawnee Station is indicative of a large portion of boilers in TVA's and the nation's utility operating base. Micronized coal reburn technology compares favorably with other NO, control technologies and yet offers additional performance benefits.

INTRODUCTION AND PROGRAM OVERVIEW

The reburn NO_x control process is essentially a post combustion cleanup technology that occurs within the boiler furnace. In reburning, NO_x is chemically reduced back to nitrogen and oxygen. Micronized coal reburn technology has application to cyclone-fired, turbo-fired, wall-fired and tangentially-fired pulverized coal units. A key advantage of micronized coal reburn, of course, is the fact that the technology uses the in-place fuel (coal) and does not require additional fuels to be brought into the plant. Research has shown that micronized fuel can function with very nearly the same effectiveness as natural gas in a reburning situation. The high effectiveness is due to the high surface area of the fuel which is conducive to the liberation of hydrocarbons and the formation of carbon monoxide; key species involved in the reburning process.

Additionally, the high burning rate of micronized coal in reburning indicates that a low tendency for carbon carryover exists; thereby creating a reduced possibility of increases in deposition in the upper radiant sections or the initial convective sections of the boiler. A major aspect of this program is to demonstrate this high effectiveness of micronized coal reburning on a full scale basis.

The reburn/OFA system can also be easily adapted to incorporate in-furnace sorbent injection for SO₂ control.

One key area of application of micronized coal reburning is for older units. Older fossil plants typically have the following operating characteristics, and many of these conditions lead to high NO_x production.

- High excess air.
- Deteriorating coal fineness.
- Poor control of secondary air.
- Mill limited from coal switching.
- Poor turn-down ratio.
- Cyclic duty operation.

TVA and many other utilities have a high population of boilers which fall into this category; yet demand upon this existing generating capacity continues. Therefore, means of reducing NO_x, which are now required under new Amendments of the Clean Air Act, while improving overall boiler performance and operability are required.

In-situ combustion modification/tuning and many types of modern low NO_x burners are suitable technologies for reducing NO_x on this class of boilers. In many cases, however, the use of these techniques will require significant upgrades of pulverization equipment, means of air distribution, and improved control systems to attain the benefits of these technologies. This significantly increases the effective cost of the technology. One key goal of this program is to demonstrate the effectiveness of micronized coal (80 percent less than 325 mesh) combined with an advanced coal returning technology to reduce NO_x without significant changes to the current firing and control equipment.

Up to 30 percent of the total fuel fired in the furnace will be micronized coal. This fuel will be injected into the upper furnace, creating a fuel-rich zone at a stoichiometry of 0.8 to 0.9. The program will examine the use of either air or recirculated flue gas as the micronized coal transport media. Overfire air will be injected at conditions for good furnace gas mixing above the reburn zone creating an overall furnace stoichiometry of 1.15 (excess air of 15 percent) and therefore change of overall boiler combustion efficiency. Cold flow modeling and numerical modeling will be used to define the parameters associated with the "best" mixing scheme for the micronized coal reburn media and the OFA.

The availability of the reburn fuel presents the potential to solve several additional problems associated with older boilers. Firstly, these units are called into deep cycling operation as they move further down the loading hierarchy. Attainment of significantly low loads (high turndown) has been restricted by low steam temperatures. With operation of the reburn injectors as true burners at low loads, steam temperatures can be better controlled. Thus, one further goal is to demonstrate the technology of operating the reburn injectors (operation at minimal air flow) as true burners (15% excess air flow) at low boiler loads for improvement of steam temperatures.

Additionally, the use of high moisture low sulfur fuels on this class of older boilers can obviate the need for installing expensive flue gas desulfurization equipment. With use of these fuels, many units will likely be faced with mill throughput limitations due to the

reduced heating value of these fuels and will incur significant generation reductions. The use of micronized coal reburning can provide the additional mill capacity needed to regain potentially lost generation capability without upgrading of the entire current mill system. Increased fuel flexibility is accomplished while, at the same time, controlling NO_x which is now a requirement for nearly all boilers.

SITE DESCRIPTION

Boilers

The host site will be one of Units 1-9 at TVA's Shawnee Fossil Plant which was built to help meet the huge electric power requirements of a nearby DOE facility. Construction began in January 1951 and was completed in 1956.

Units 1-9 are 175 MWe (gross) front wall-fired, dry-bottom furnaces burning East Appalachian lower-sulfur coal. The plant was originally designed to burn high-sulfur coal; but in the 1970's, the plant was modified to burn low-sulfur coal in order to meet an emission limit of 1.2 lbs SO₂/10° Btu of heat input without the use of any sulfur dioxide control technology. Each unit has been equipped with a baghouse to control particulate emissions. Flue gas from each unit discharges to one of two 800-foot stacks, also constructed in the 1970's. The nine existing pulverized coal units are representative of a large number of wall-fired units in the industry which will be required to reduce NO_x emissions in response to the 1990 Clear Air Act Amendments. Unit #6 has been selected as the demonstration unit. This unit is identical to the others with the exception that different burner air registers have been installed.

Coal Acquisition

TVA has contracts in place to supply Shawnee with low-sulfur bituminous coals from Kentucky and West Virginia. These coals will be used as the primary fuels for the project. TVA has test burned western coals such as Powder River Basin (PRB) at a number of sites, including Shawnee, since the late 1970's. PRB coal will be obtained for testing during this demonstration.

MICRONIZED COAL TECHNOLOGY

Technology Description

The technology to be utilized is a combination of a technology that produces microfine coal reliably and economically, with new applications of a relatively well known NO_x control technology (fuel reburning). When micronized coal is fired at a stoichiometry of 0.8 to 1.2, devolatilization and carbon conversion occur rapidly.

Micronized coal is defined as a coal ground to a particle size of 80 percent 43 microns or smaller. The MicroFuel® system, consisting of the MicroMill and an external classifier, micronizes coal to a particle range of 10 to 20 microns.

The combined surface area of just one gram of micronized coal particles is 31 square meters, contrasted to a surface area of 25 square meters per gram for pulverized coal.

The MicroMill system is a patented centrifugal-pneumatic mill with the replacement rotating impeller as the only moving part. Size reduction is accomplished by the particles themselves striking against one another as they whirl in a tornado-like column of air inside the MicroMill. Centrifugal force retains material in the cone and rotational impact zone (RIZ) as the particles reduce in size prior to being conveyed by the air stream entering the center of the rotating impeller.

The net result of micronized coal as a reburn fuel is a uniform compact combustion envelope allowing for complete combustion of the coal/air mixture in a smaller volume than conventional pulverized coal. Heat rate, heat flux, carbon loss, and NO_x formation are all impacted by coal fineness.

DESIGN OF THE MICRONIZED COAL REBURN/OFA SYSTEM

Design Parameters

The success of this demonstration hinges on designing a reburn system that will rapidly mix the micronized fuel with crossflowing flue gas rapidly and as completely as possible. The key parameters to mixing in this situation are fuel inlet velocity, inlet area and number of inlets. With a properly designed mixing system most or all of the flue gases will pass through a region of controlled fuel rich stoichiometry. Generally, a region of less than 0.9 of stoichiometric is required. Gases not exposed to the reburn media or hydrocarbons liberated from the reburn media will remain untreated. Thus, with incomplete mixing, NO_x reduction effectiveness is compromised. A design criteria of >70% of the main combustion zone flue gas mass flow through the fuel rich environment will be employed.

Additionally, with incomplete mixing of the reburn media super-fuel rich region can be generated wherein the hydrocarbon products are not utilized by the NO_x in the flue gas. In this situation, hydrocarbon emissions can increase, particles can coke and become very difficult to burn and tendencies for increased tube deposition are possible. The OFA system can compensate for fuel unmixedness in some cases but not likely for extreme situations. Thus, an additional goal is to not have large super-fuel rich regions.

In addition to mixing, establishing sufficient residence time for the NO_x reduction reactions to occur is necessary. Conversion effectiveness varies with residence time. A design criteria for residence time to be >0.4 seconds for high reduction will be employed. The residence time requirement dictates the vertical location of the fuel inlets. Flue gas temperature also plays a role in reburn residence time requirements and reburn fuel injector location.

Similarly, the attainment of high mixedness in the OFA system is important. The OFA system completes the oxidation of the fuel rich flue gases. Increases in unburnt carbon and increasing tendency for slag deposition can occur as a result of OFA unmixedness. A design criteria of 99% mixing is required to satisfactorily complete combustion in this zone. Note also that mixing must be controlled so as to avoid creation of high temperatures and air rich regions which can regenerate NO_x.

The same degree of importance is placed on residence time in this zone. A design criteria of residence time > .3 seconds will be employed in the post reburn zone.

Boiler performance factors such as furnace exit temperature, the distribution of furnace exit temperature, and boiler heat flux profiles are other criteria of importance to be considered in the design of the reburn/OFA system.

The key design parameters to be established are:

Reburn Injection

total quantity location velocity

number

transport media (air or FGR)

OFA Injection

total quantity location velocity number

Boiler Performance

peak heat flux vertical heat flux profile furnace exit temperature spatial furnace exit temperature variation

Boiler Reliability

avoid fuel rich regions on walls avoid fuel rich regions at exit

Design Process

The combustion process in a coal fired boiler is a complex process. High quantities of chemical energy are converted into heat in a relatively small volume (short time). The generation, transfer, and transport of this heat involves high intensity turbulent processes. As a result, the velocities, temperatures, and gas compositions exiting the furnace region of a boiler deviate significantly from a simple plug flow scenario. To account for the interactions

between a complex ill defined flow and provide the required confidence of mixing first fuel then air very nearly completely, requires several approaches.

To meet the target design criteria and establish high degree of confidence in the performance of the reburn/OFA design several resources will be employed. These include:

numerical modeling of the flow and mixing processes physical flow modeling good boiler test data with furnace probing experience in mixing and fluid mechanics experience in fundamental combustion processes research and full scale data from other similar programs

Physical flow modeling will be conducted utilizing dynamic similarity in plastic models. Smoke and other chemical tracers will be used to assess mixing profiles and establish velocities.

QA comprehensive boiler and furnace test program will be conducted. Furnace temperatures will be established at different locations for a range of boiler conditions. These will be used to verify the numerical model. This data will be used in addition to develop preliminary designs that will be evaluated in more detail and refined by the physical and numerical modeling efforts.

Numerical modeling will be carried out by adapting the Radian Furnace Simulation Model (FSM) to the Shawnee #6 configuration and incorporating reburn and OFA inlets.

USE OF NUMERICAL MODELING IN THE DESIGN PROCESS

The Radian FSM

The Furnace Simulation Model (FSM) was developed to provide assistance in determining how a particular burner and/or burner overfire air system will perform in a given boiler. It is a complete model of the combustion, fluid mechanics, and heat transfer processes occurring in the boiler. The model provides the ability to analytically change burner designs, add OFA ports, and move burners around. The model predicts NO_x levels but more importantly provides an evaluation of the potential for operational problems (i.e., slagging, corrosion, performance, heat transfer maldistribution) for a particular burner type and/or burner overfire air system configuration. The model provides the ability to "look" inside a boiler with the purpose of diagnosing problems where measurements are difficult or impossible. Trained application of the model permits evaluation of complex tradeoffs between NO_x control techniques and operational benefits and penalties.

The FSM model is a complete two-phase simulation of the combustion, fluid mechanics, and heat transfer process occurring in a boiler furnace. It is designed to run within the PHOENICS Navier-Stokes equation solver and is capable of incorporating the detailed geometry of each burner. In addition, a complete description of the walls and heat

absorbing surfaces within the boiler is used by the model to incorporate the flow resistance and heat transfer effects of the walls. Operation specific fuel properties are also used. Thermal and fuel nitrogen chemical kinetics are included on a simplified basis that incorporates experience factors.

The model includes fuel devolatilization, gas phase combustion, heterogeneous combustion, and interphase transport processes of heat, mass, and momentum. Radiant heat transfer between particles, gas, and walls is included using a six flux relationship for each phase. Convective heat transfer is included based upon relationships known to be useful in furnace design. Turbulence at the microscale, eddy scale, and in large recirculation zones is also included.

The model is run using a unique three pass approach which saves computer costs and time. The first pass is a coarse combustion analysis which solves for the major combustion species along with fluid mechanics and heat transfer. In the second pass, the fluid mechanics and heat transfer are fixed and dissociation of species to form free radicals such as O and N atoms are included based upon chemical equilibrium calculations. These calculations are performed at each grid point and a Gibbs free energy minimization approach is utilized to adjust species concentrations and consider dissociation. The NO_x kinetics are evaluated in the third pass, again with the fluid mechanics fixed. This approach reduces the number of solved parameters (and subsequent computer time) in terms of the number of species for the main run, where their influence on the fluid mechanics solution is minimal.

The furnace geometry is divided into grids in three dimensions. A fine grid is used in the burner region where gradients of concentration, temperature, and velocity are high while a much coarser grid is used in regions of the furnace outside the main combustion region. The current model solves for 18 field variables and incorporates about 18,000 grid points for the initial geometric configuration.

Adaptation of the Numerical Model

Initial work in adapting the model involves the gathering of information necessary for input and verification of the model. Primary data include the current burner/boiler geometry information in the form of drawings and sketches. The model requires details of the burners sufficient to determine the flow areas and velocities in each region (primary, secondary, and tertiary). Estimations are made of tangential (swirl) velocities based upon drawings and flame observation. These estimations are used for initial runs, and are parameters that can be varied during the course of the project. Also required is the shape of the water tubes around the throat.

For the boiler, the required data includes the general geometric arrangement and design values for the water wall conditions (i.e., temperature versus height or depth) for each plane or surface. The location and design temperatures of division walls, partial water walls, and any other surface extensions are also required.

Typical fuel analyses (ultimate and proximate) are required. Ash fusion temperature data is also desirable, as are results from any recent pulverization tests and any recent data on fuel/air balancing tests.

Any good NO_x data relating NO_x to load, excess oxygen, and fuel types is also required for verification. Available information on the occurrence of operational problems such as slagging, tube corrosion, heat rate, etc., is also be collected from operating plant personnel to assist in the verification work. In this demonstration program comprehensive tests of furnace conditions will be performed to assist in the verification.

In the initial runs of the model work is performed to optimize the grid size and grid size distribution to be the coarsest that will give good results, and yet have acceptable run times (and corresponding computer run time costs). Key to the description of combustion/NO_x processes is having a sufficiently fine grid in the near burner region where the gradients of temperatures, velocities, and compositions are the highest. In the bulk furnace regions of the model, the grid can be much coarser and still describe heat transfer and mixing processes adequately. Simply developing the entire grid as fine as that required in the burner region would result in extremely long run times. (e.g., perhaps as long as a week on a high speed 486 PC for a single run.) The results of this task are examined for tradeoffs relative to a baseline set of data.

In the actual model verification runs the objective is to verify that the model accurately describes the NO_x formation rates and operational characteristics of the current boiler configuration. This work is divided into the following three areas.

With a uniform "ideal" firing pattern for each burner, the model is run at conditions of load, excess oxygen, mills-out-of-service, etc., for which good NO_x and operational data exist. Once a good comparison in trends and levels is achieved, work proceeds. Key process constants are then adjusted on an as needed basis to improve the predictability of the model. These results are also examined for trends consistent with frequently encountered operating problems.

In this work any suspected fuel or air imbalances are incorporated into the model. Maldistribution of fuel and primary air are simple changes in input variables to the model. In addition, any broken registers, bad pipes, or other combustion anomalies are also included.

Cases are then run to compare the "real world" operation with the NO_x and operational problems documented previously. Parametric variations are then made around these initial values to determine the sensitivity of NO_x and operational characteristics to firing anomalies. Comparisons are also made with the previously developed process constants to determine if any additional adjustments are necessary.

With a well calibrated model, the design of the reburn OFA system proceeds. Additional fuel injectors are incorporated into the model as are OFA ports. Parametric variations are made of injection flow, location, velocity, number in inlets, and the benefits of tilt, yaw, and swirl.

SYSTEMS EVALUATION USING THE FSM.

A key use of the model will be to evaluate in extensive detail the performance of the overall system once the critical design parameters have been established. The model will be used to:

Evaluate Low Load / Peak Load Burner Performance

Examining Potential Critical Localized Problems

- Coal Pipe Temperatures
- Burner Component Temperatures
- Slag Deposition in Reburn or OFA Regions

Determining Critical Control Measurements and Ranges of Values

- Excess Oxygen Levels
- Main Fuel Flows
- Reburn Fuel Flows

Firing Procedures During Micronizer Outages

Firing Procedures for Alternate Coal (L.S. PRB)

A UNIQUE FUEL INJECTOR/BURNER IS REQUIRED

Much attention in this program will be given to the design of the reburn injector/burner. The functions of this burner are unique to most of the common requirements of the burner industry.

In the reburn mode, the burner will be designed to utilize as little as possible secondary air. Any significant amounts of secondary air increase the amount of reburn media required. The secondary air flow will be only that required for cooling of the burner mechanism. In the reburn mode, consideration will be given to the use of recirculated flue gas as the transport media for the micronized coal. This will further reduce the required amount of reburn media.

In low load operation of the boiler, however, the fuel injector will be designed to function as a stand alone burner. This requires about 9:1 lbs air/lb coal. Thus the secondary air will be required to modulate over an order of magnitude range.

With use of low sulfur PRB coals, it is likely that the main pulverizers of the unit will encounter a significant throughput derate. In this situation the micronized coal system will be required to carry a significant percentage of the boiler fuel input. Perhaps, as high as 30

percent. Thus, significant fuel modulation over about a 10:1 range is also required for the specially designed injector/burner.

The design of a modulating burner that will vary both fuel flow and air flow over a wide range of values presents a unique design challenge in this program.

PROGRAM SCHEDULE

The program schedule is outlined as followed:

Detailed Testing

Modeling

Completed December 1993

Completed February 1994

Final Design

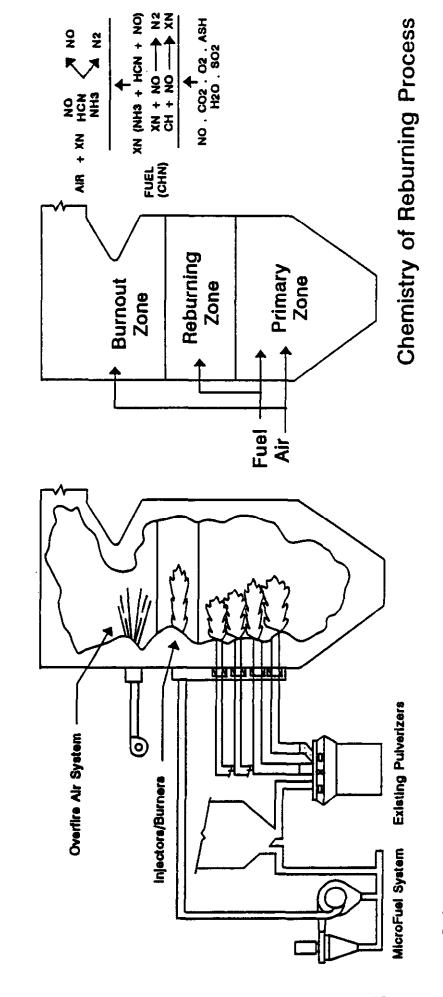
Completed May 1994

Completed March 1995

Final Testing and Report

Completed July 1995

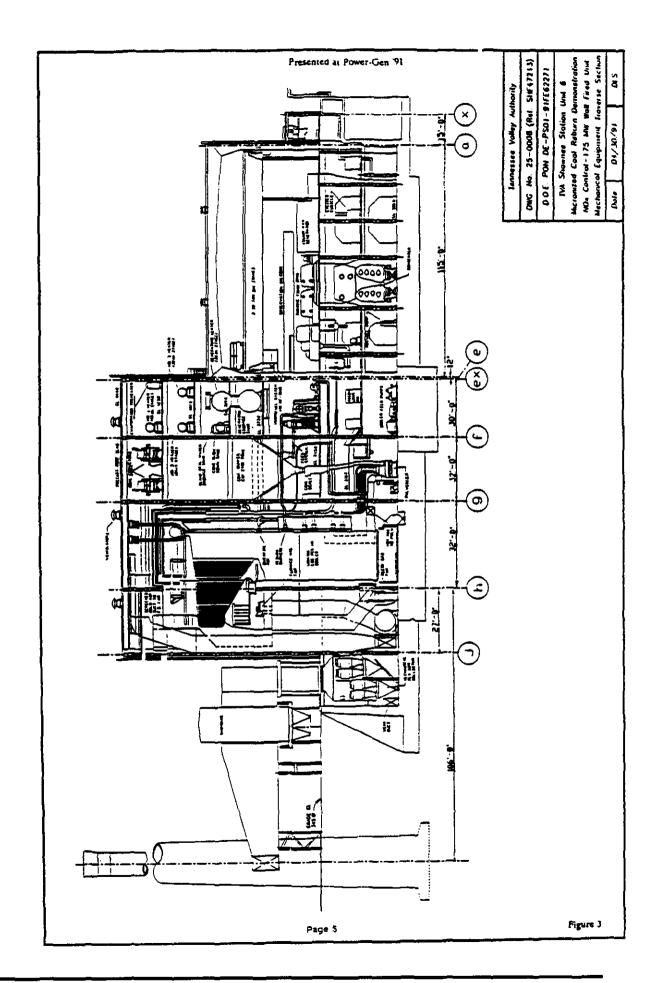




Schematic of Reburn Process

PROGRAM BACKGROUND

- Micronized Coal Reburn High Candidate System
 - Cyclone Fired Systems
 - Turbofire Fired System
 - PC Systems
- Micronized Coal Reburn Attractive Alternative for Older Boilers
 - Imprecise Control of Air Flow and Distribution
 - Deteriorating Fineness and Flow Maldistribution
 - Marginal Milling Capacity for Low Sulfur Fuels
- Micronized Coal Reburn Attractive for Deep-Cycling Boilers
 - Raise Low Load Steam Temperatures
 - Provide Peaking

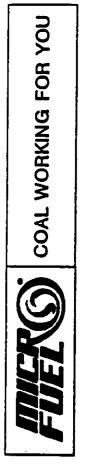


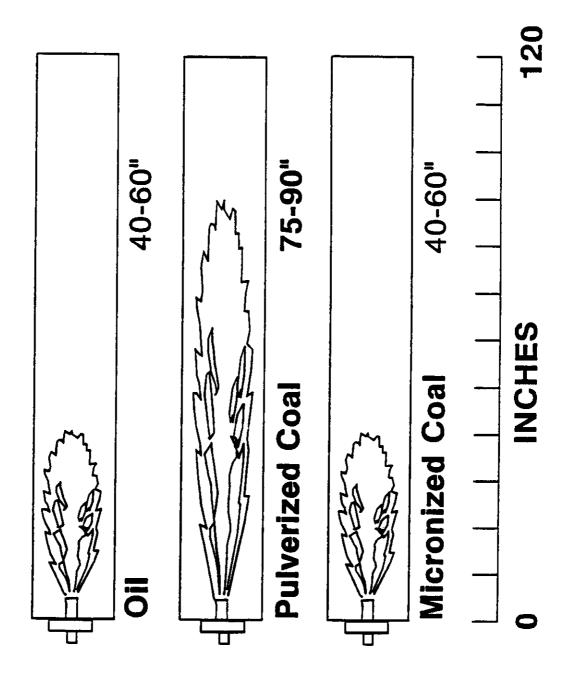
MICRONIZED COAL AS REBURN MEDIA

- Use Same Fuel as Main Combustion Zone
- High Surface Area Liberates Hydrocarbons
- Rapid Burnout to Avoid Convective Pass Deposits
- With Good Mixing Can Require Less Reburn Media



Micronized Coal Flame Characteristics





GOALS OF DEMONSTRATION PROJECT

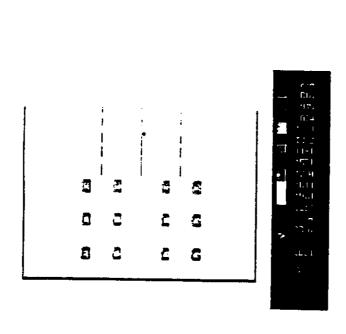
- Full Scale Demonstration of Micronized Coal Reburn
- Use of Micronized Coal Reburn to Improve Turndown / Peaking Capabilities
- Show Improved / Non-Undegraded Boiler Performance
- Demonstrate Micronized Coal Reburn Injectors Operate As Burners
- Use Micronized Coal Reburn to Increase Fuel Flexibility

RESOURCES EMPLOYED IN REBURN / OF A SYSTEM DESIGN

- Numerical Modeling
- Cold Flow Modeling
- Burner / Mixing / Fluid Mechanics Experience
- Combustion Expertise
- Review of Other Projects

RADIAN FURNACE SIMULATION MODEL

- Computational Fluid Dynamics
- Full 3-Dimensional
- Two Phase Flow
- Radiant and Convective Heat Transfer
- Interphase Transport
- Well Calibrated



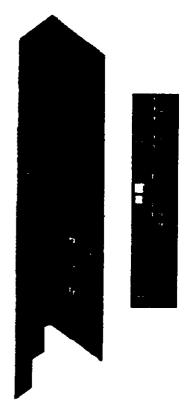
Burner Configuration



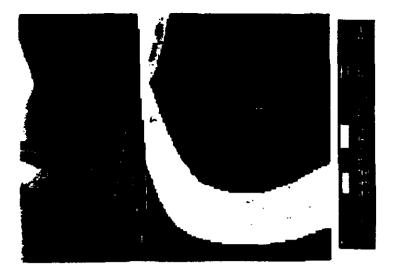
Gas Temperature Near Sidewall and Firing Face



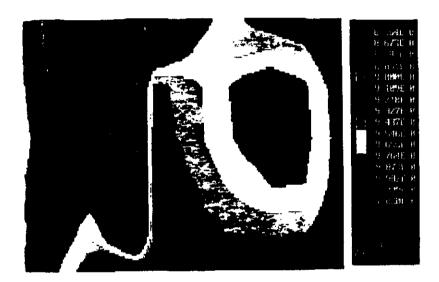
Gas Temperature ল প্রয়ালন Region



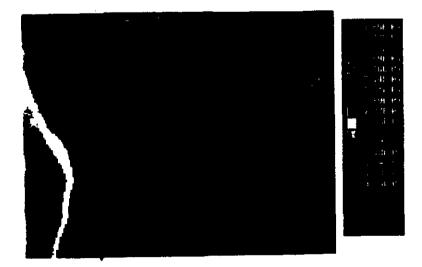
Equivalence Ratios Near Sidewall and



Exit Plane Gas Femperatures



Exit Plane Equivalence Ratio



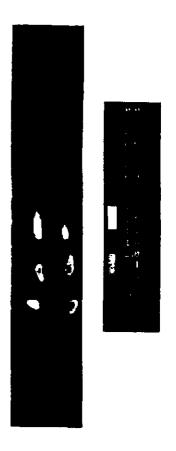
NOx Concentrations at Exit



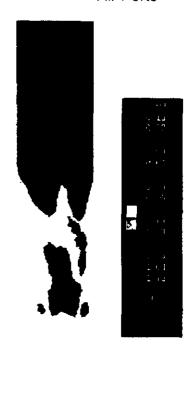
_arge Overfire Air Ports



Small Overfire Air Ports



Poor Mixing from Large Overfire Air Ports



Good Mixing from Small Overfire Air Ports

MODELLING IN REBURN SYSTEM DESIGN

- Design "Best" Reburn / OFA System
- Set Reburn System Parameters
 - Fuel Quantity
 - Injection Velocity
 - Location of Inlets
 - Number of Inlets
 - Evaluation Benefits of Tilt, Yaw, Swirl
 - Use of FGR as Coal Transport
- Set OFA System Parameters
 - Air Quantity
 - Injection Velocity
 - Number of Inlets
 - Evaluate Benefits of Tilt, Yaw, Swirl

COMBUSTION / NO, CRITERIA OF 'BEST' DESIGN

- Reburn Mixed Mass (Mass of Flue Gas < 0.9 stoich)
- Residence Time Criteria (Mass of Flue Gas < 0.9 stoich with $T > T_{REO.}$)
- OFA Mixed Mass
- OFA Residence Time

BOILER PERFORMANCE CRITERIA OF "BEST" DESIGN

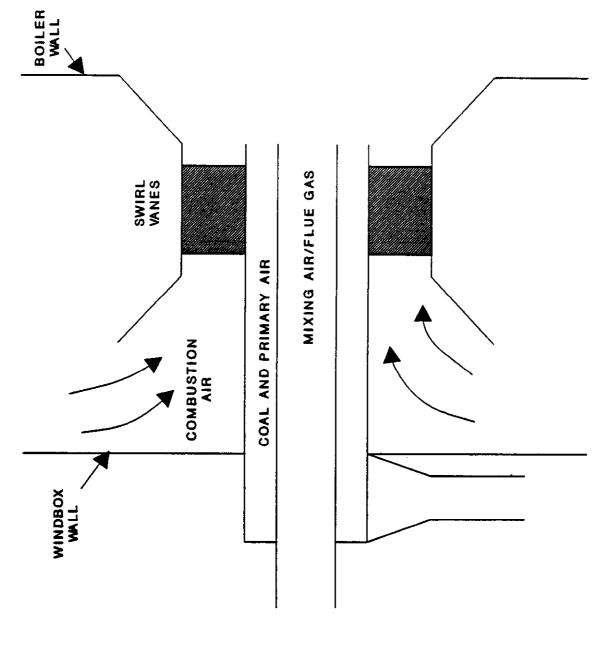
- No Change from Baseline
 - Average Furnace Exit Temperature
 - Spatial Distribution of Furnace Exit Temperature
 - Wall Vertical Heat Flux Profile
 - Peak Wall Heat Flux
 - Mass Flow Velocity Distribution
 - Carbon Burnout
- Avoid Large Fuel Rich Regions on Walls
- Avoid Large Fuel Rich Regions at Exit Plane

MODEL ADAPTATION TO SHAWNEE #6

- Incorporate Boiler / Burner Geometry
- Incorporate Current and Planned Fuel Properties
- Boiler Characterization Testing Program
 - Gas Temperatures
 - Gas Velocities
 - Species Concentrations
- Review / Correlate Unit Operational Data
 - Carbon Burnout
 - Steam / Metal Temperatures
 - Tube Wastage
 - Unknown or Suspected Fuel / Air Imbalances
- Review Correlate Unit Emissions Data
 - NO,
 - CO
 - Excess Oxygen Levels
- Verification of Model with Tests / Operational Data
 - Perform Verification Runs
 - Correlate Results

USE OF MODEL IN SYSTEM PERFORMANCE EVALUATIONS

- Evaluate Low Load / Peak Load Burner Performance
- Examining Potential Critical Localized Problems
 - Coal Pipe Temperatures
 - Burner Component Temperatures
 - Slag Deposition in Reburn or OFA Regions
- Determining Critical Control Measurements and Ranges of Values
 - Excess Oxygen Levels
 - Main Fuel Flows
 - Reburn Fuel Flows
- Firing Procedures During Mill Outages
- Firing Procedures During Micronizer Outages
- Firing Procedures for Alternate Coal (L.S. PRB)



Session 2 Advanced Electric Power Generation Systems

Co-Chairs:

Larry K. Carpenter,

Morgantown Energy Technology Center/ U.S. Department of Energy **George Lynch**,

Office of Clean Coal Technology/ U.S. Department of Energy



YORK COUNTY ENERGY PARTNERS ACFB DEMONSTRATION PROJECT STATUS

S. I. Wang and F. T. Bolinsky
Environmental and Energy Systems
Air Products & Chemicals, Inc.

Allentown, Pennsylvania

ABSTRACT

The York County Energy Partner, L.P. project, to be located in York County, Pennsylvania, will demonstrate the world's largest atmospheric circulating fluidized bed (ACFB) boiler under sponsorship of the U.S. Department of Energy's Clean Coal Technology I Program. The single ACFB boiler, designed by Foster Wheeler Energy Corporation, will supply 227 MWe of net electrical power and export up to 400,000 lb/hr of steam to an adjacent paper mill. This paper outlines the project summary, process description, changes due to site relocation, the value improvement of boiler island and current status of the project.

INTRODUCTION

The York County Energy Partners cogeneration project located in York County, PA will demonstrate the largest single ACFB boiler in the U.S. under sponsorship of the U.S. Department of Energy's (DOE) Clean Coal Technology I Program. The goal of the DOE program is to demonstrate the technical and economic feasibility of applying circulating fluidized bed combustion technology at the 250 MW scale for producing electrical power and steam in an environmentally acceptable manner while efficiently utilizing our nation's coal resources. The single-train ACFB boiler, designed by Foster Wheeler Energy Corporation (FWEC), will supply 227 MWe of electrical power to the Metropolitan Edison Company (Met-Ed) and export approximately 400,000 lb/hr of

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back-up the YCEP facility when it is shutdown for maintenance) and provide YCEP significant emission offsets. An additional significant site change item is the water supply. The YCEP facility will use treated wastewater from Glatfelter's secondary water treatment facility (3 million gpd). This largely eliminates the facility's needs for fresh water as compared to the previous site. Most significantly, this project will result in a decrease in SO₂ emissions of approximately 50%. Table 1 compares the scope differences between the old site and new site.

BOILER MODIFICATIONS

As a result of increasing export steam from 40,000 lbs/hr (15 psig) to 400,000 lbs/hr (600 psig, 680°F), the boiler steaming rate will increase from 1,725,000 lbs/hr to 2,100,000 lbs/hr. The combustor size increases proportionally. Table 2 compares the key boiler parameter differences between the two sites. As more pilot plant data becomes available from Foster Wheeler's pilot plant testing, Foster Wheeler is confident that they can reduce the cyclone diameter from 21 ft. to 20 ft. in I.D. and reduce the front coal feeders from 8 to 6, without sacrificing the combustion efficiency and emissions performance. The boiler configuration is shown in Figure 3.

VALUE IMPROVEMENT

The pilot plant test results conducted by Foster Wheeler suggested that the SNCR process (NH₃ thermal deNOx) can further reduce NOx emission from 0.15 to 0.10 lb/MM BTU. Furthermore, the SOx removal efficiency can maintain at 92% at a lower Ca/S ratio. These data are presented in Tables 3 and 4. More pilot plant testings are scheduled to further optimize the thermal deNOx process using different reagents.

PROJECT STATUS

The commercial activity and status is shown in Table 5. The Public Utility Commission approved the power contract in May 1993. In June 1993, YCEP and DOE executed a

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modification to the Cooperatiave Agreement for the North Codorus site. An Agreement for steam supply to P. H. Glatfelter is currently being negotiated. Other commercial agreements, such as the coal contract, limestone contract, and NOx offset agreement, and ash byproduct utilization agreements, are in progress. By October local land development approvals are expected to be in hand.

As far as project schedule is concerned, we are continuing to work on environmental permitting, equipment procurement and preliminary engineering site work. Construction is expected to start in January 1995. Boiler erection will start in September 1995 and the boiler island and turbine generator erection will be completed in September 1997. The first fire is scheduled in September 1997 and the commercial operation is scheduled for December 1997. This is shown in Tables 6 and 7.

YCEP Project Summary

Title:

York County Energy Partners Clean Coal Technology Round

Cogeneration Project

Proposer:

York County Energy Partners, L.P., a Project Company of

Air Products and Chemicals, Inc.

Location:

Technology:

York County, PA

Atmospheric Circulating Fluidized Bed

Combustion

Applications:

Utility and Industrial Electric Power/Steam Generation,

Repowering Existing Boilers or New Plants

YCEP Project Summary

Fuel:

Less than 2% Sulfur Bituminous Coal

Size:

2,100,000 PPH/2500 psig/1005°F 227 MWe net to Met-Ed

Main Steam, 1,325,000 PPH/495

psig/1005°F Reheat Steam

P.H. Glatfelter Co., Spring Grove, PA

Greater than \$380 Million

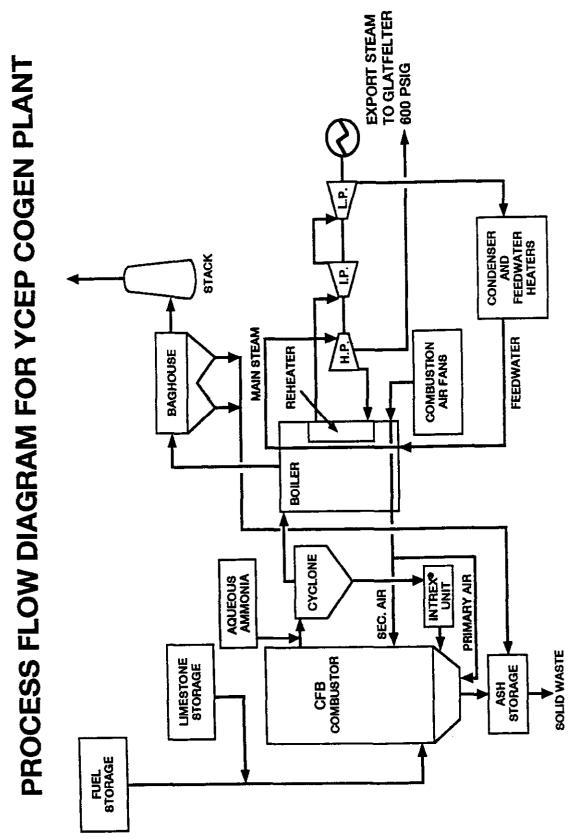
400,000 PPH Steam

\$75 Million

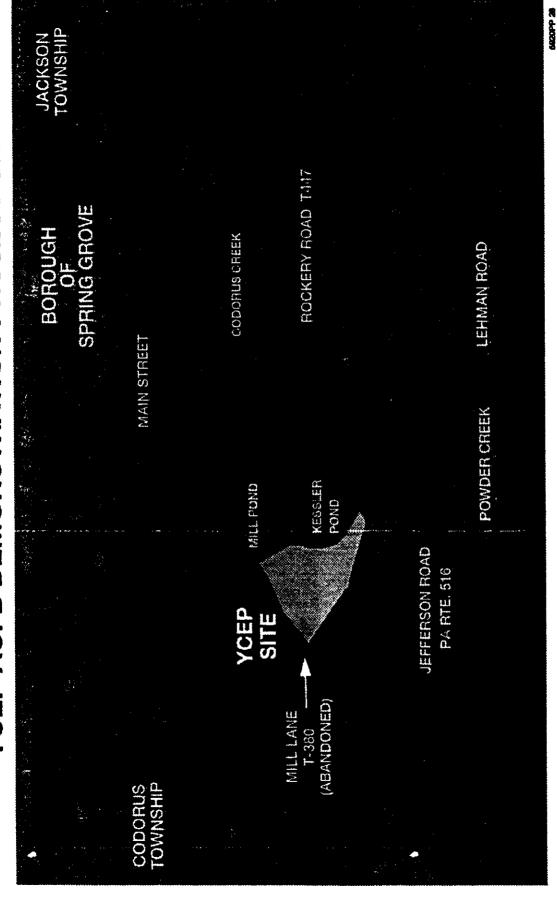
DOE Funding:

Project Cost:

Steam Host:



YCEP ACFB DEMONSTRATION PROJECT SITE



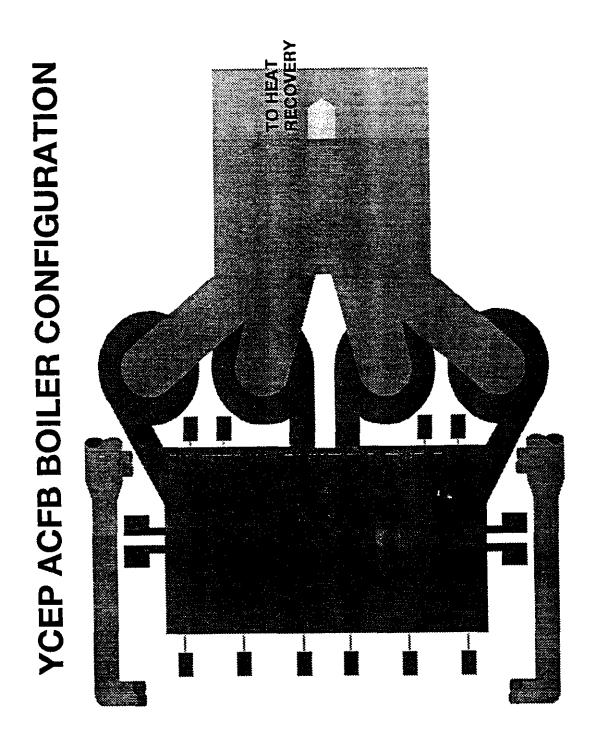
PROJECT SCOPE COMPARISON FOR YORK COUNTY ENERGY PARTNER ACFB DEMONSTRATION PROJECT

Old Site	Project Scope	New Site
West Manchester Twp.	Site Location	North Codorus Twp.
J. E. Baker, Refractory	Steam Host	P. H. Glatfelter Paper Manufacturer
Dilon Mailulactures	Site Size (Acres)	25
766	Electricity (MW) to Met-Ed	227
1,725,000	Total Steam Made (lbs/hr) 2,500 psig, 1005°F	2,100,000
1.400.000	Reheat Steam (lbs/hr) 1005°F	1,400,000
40.000 (15 psia)	Export Steam (lbs/hr)	400,000 (600 psig, 680°F)
Yes	New Auxiliary Boiler	No
City Water	Cooling Water Supply	P.H.G. Co. Secondary Effluent From Mills
High Bearing Capacity 15 ft Deep Over Burden	Soil Conditions	Low Bearing Capacity 60-100 ft Deep Over Burden
Double Circuit 230 KV 1.25 Mile to Jackson Substation	Electric Interconnection	 Double Circuit 115 KV 0.5 mile on P.H.G. Property Single 115 KV, Nearby Met-Ed Interconnection Point

5020PP 28

KEY BOILER PARAMETERS

New Olic		
2,100,000	Main Stream, lb/hr	1,725,000
1,400,000	Reheat Steam, Ib/hr	1,400,000
23	Depth, ft	20
84	Width, ft	78
105	Height, ft	112
20	Cyclone Diameter, ft	21
4	No. of Cyclones	4
	Coal Feeders:	
9	- Front	8
4	- Back	4



EMISSIONS IMPROVEMENT

Old Design	0.15	%76	0.15	0.004
	NOx, Ib/MMBtu (SNCR)	SO ₂ Removal Efficiency	CO, Ib/MMBtu	VOC, Ib/MMBtu
New Design	0.10	95%	0.15	0.004

EFFICIENCY IMPROVEMENT

Old Design	1,952	hr 1,988,700	2,152,000	88.19	170,296	92	2.5	28,788
	Steam Duty, MMBtu/hr	Total Combustion Air, lb/hr	Total Flue Gas, lb/hr	Overall Efficiency, %	Fuel Flow, lb/hr	Percent S Removal	Ca/S Ratio	Limestone Flow, Ib/hr
Revised Design	2,290	2,299,300	2,486,000	89.21	197,076	92	2.3	32,940

YORK COUNTY ENERGY PARTNERS **COMMERCIAL STATUS**

- Met-Ed Power Contract
- P. H. Glatfelter Steam Agreement
- Ash Byproduct Beneficial Use Agreement
- · Coal Contracts
- Limestone Procurement
- NOx Offset Agreement
- Local Land Development Approvals

YCEP ACFB DEMONSTRATION PROJECT SCHEDULE

P. H. GLATFELTER SITE

plete
s Com
roces
PA P
Z

Jan. 95 Jan. 95

Sept. 95

Begin Boiler Steel Erection

PA DER Review Complete

May 97

Sept. 96 - Sept. 97

Erect Turbine/Generator

Commission DCS

First Fire

Aug. 97

Sept. 97

Dec. 97

97 Dec.

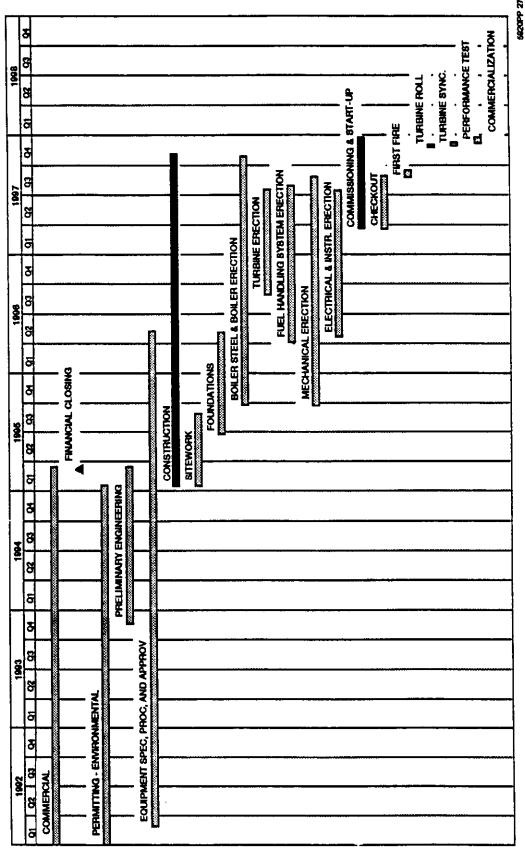
T/G Load/Sync

Commercialization

Hydro Boiler

YORK COUNTY ENERGY PARTNERS COGENERATION PROJECT

Overview Schedule: 6 July 1993, 12:25p.m.; S. J. McKitish, P.E.





DMEC-1 Pressurized Circulating Fluidized Bed Demonstration Project

Gary Kruempel and Steve Ambrose
Midwest Power
Des Moines, Iowa

Steve Provol
Pyropower Corporation
San Diego, California

Mitch Bjeldanes Black & Veatch Kansas City, Missouri

Second Annual Clean Coal Technology Conference

Sponsored by DOE
Southern States Energy Board
Atlanta, Georgia

September 7-9, 1993

DMEC-1 Pressurized Circulating Fluidized Bed Demonstration Project

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Steve Provol
Pyropower Corporation
P.O. Box 85480
8925 Rehco Road
San Diego, California 92186-5480

Mitch Bjeldanes Black & Veatch 8400 Ward Parkway P.O. Box 8405 Kansas City, Missouri 64114

INTRODUCTION

The Des Moines Energy Center (DMEC) Project will be the first commercial scale demonstration of Pyropower Corporation's PYROFLOW® Pressurized Circulating Fluidized Bed (PCFB) technology. The project will be a repowering of an existing steam turbine at the DMEC site. The design incorporates a hot (1,600° F) particulate removal system and operates in a combined cycle configuration for increased plant efficiency.

The DMEC-1 limited partnership, with Dairyland Power as the limited partner and Midwest Power, formerly Iowa Power, as the general partner, will be the participant for the project. The project was selected in the Clean Coal Technology Round 3 solicitation. The partnership signed the Cooperative Agreement with the DOE in May 1991.

In August 1991, Midwest Power, Dairyland Power Cooperative, Pyropower Corporation, and Black & Veatch initiated the preliminary design of the PCFB Repowering Project. During

the preliminary design process, plant and system layouts have been completed, subsystem specifications have been prepared, and cost and schedule baselines have been updated. Process verification testing for hot gas filter equipment, gas turbine materials, and fuel selection has continued at the Ahlstrom PCFB Testing Facility in Karhula, Finland. Testing results have shown the need to continue this testing prior to finalization of the ceramic filter system selection.

PROJECT GOALS

The goals of the project are to demonstrate the following advantages of the PCFB technology:

- Lower capital cost compared to atmospheric CFB or pulverized coal plant with scrubbers.
- High efficiency and reduced CO₂ emissions.
- Reduced space requirements.
- Hot gas cleanup technology.
- No exposed surfaces in the lower combustor.
- Control of NO_x, SO_x, and CO.
- Simplified load following.
- Erosion prevention.

DMEC-1 PROJECT SITE DESCRIPTION

The Des Moines Energy Center (DMEC) is located southeast of the City of Des Moines, Iowa, in the City of Pleasant Hill. The plant is located adjacent to the Des Moines River on Highway 46.

DMEC was first constructed in 1925 with the installation of two steam turbines and six stoker fired boilers. Between 1925 and 1964, five steam turbine generators and five pulverized coal fired boilers were added. The units operated in baseload mode up to the late 1970s when their operation was reduced due to the addition of more efficient

generating units to the Midwest Power system. Steam Turbine Number 6, which will be repowered in the DMEC-1 Project, was mothballed in 1985.

The DMEC-1 Project will require refurbishment or replacement of some major plant equipment. The existing turbine generator is expected to be refurbished. It is rated at a nominal 65 MWe and is a nonreheat unit designed to operate at 1,250 psig and 950° F with a steam flow of approximately 561,000 lb/h. In addition, the existing coal handling facilities, structure, and some of the major auxiliaries, such as the boiler feed pumps, condensate pumps, circulating water pumps, fuel oil, condensate and surge tanks, deaerator, feedwater heaters, auxiliary heat exchangers, etc., will most likely be refurbished with some components being replaced. New equipment is expected to include the main step-up transformer, main auxiliary and reserve auxiliary transformers, digital control system, gas turbine, electrical distribution system and switchgear, and demineralizer.

An artist's rendering of the reconstructed plant structure is shown as Figure 1.

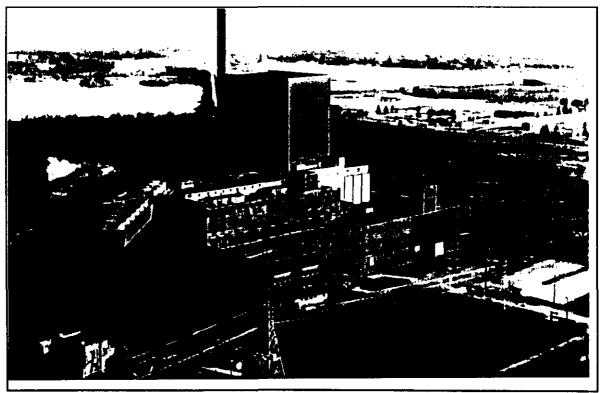


Figure 1. Artist's Rendering DMEC-1 PCFB Repowering Demonstration

TECHNOLOGY DESCRIPTION

The PCFB process uses a combined cycle which employs a combination of a gas turbine and a steam turbine to generate electrical power. The pressurized combustion chamber is used to burn coal to produce steam for the steam turbine which produces approximately 75 percent of the total plant output. The hot flue gases are filtered and expanded through a turbine to generate the remaining 25 percent of the plant output and to drive the compressor that supplies air to the PCFB combustor. A schematic process flow diagram of the Pyroflow PCFB process and subsystems is shown on Figure 2.

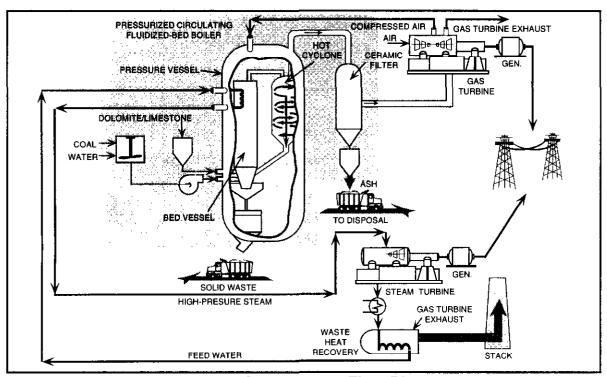


Figure 2. PCFB Process Flow Diagram

Coal and sorbent are fed to the combustor as a paste, applying existing technology as used in the pumping of concrete, and atomized with steam to distribute the materials in the PCFB. The compressor section of the gas turbine provides air to the PCFB vessel. The air flows from the top to the bottom of the vessel, and cools the vessel and internal components. The fuel and sorbent are mixed with the air in the combustor chamber where combustion occurs at about 1,600° F. Heat is recovered from the hot flue gases in a similar way as conventional boilers to generate steam to power the steam turbine.

As particles are burned and sulfur is absorbed in the furnace by the sorbent, the finer particles of coal and sorbent become entrained with the flue gas and enter the hot cyclone. Here, the majority of the particles are collected and returned to the combustor through the loop seal. The finer ash particles continue with the hot gases to the ceramic filter where final removal of particulates is achieved.

Once cleaned, the hot pressurized flue gases are expanded through the gas turbine. The resulting mechanical power drives the compressor and the gas turbine's generator. The remaining useful heat exhausted from the gas turbine is recovered in a conventional heat recovery unit to preheat process water in route to the PCFB boiler. Clean exhaust is then released to the stack.

TECHNOLOGY DEVELOPMENT HISTORY

The Karhula PCFB Testing Facility was built in Karhula, Finland, to support the design and operation of commercial first generation and advanced PCFB units. In 1989, Ahlstrom, the parent company of Pyropower, initiated operation of the Karhula PCFB facility. It is an integrated PCFB unit, including all of the key components and incorporating the same mechanical design features which will be utilized in commercial plants. These include fuel handling and preparation systems, sorbent injection systems, pressurized furnace with radiant heat transfer surfaces, hot cyclone, ceramic filter, ash cooling and depressurization systems, and testing of materials and coatings for gas turbine blades.

The main objectives of the Karhula PCFB Filter testing program are the following:

- To generate process data for the design of commercial size PCFB units.
- To develop engineering data for design of PCFB systems and plant auxiliaries including fuel feeding and ash handling.
- To generate database information for auxiliary equipment performance which can be used for other advanced coal utilization technologies.

 To demonstrate a commercial scale high-pressure high-temperature filter under PCFB conditions.

PILOT PLANT TESTING RESULTS

The facility has operated for over 3,000 hours with various sorbents and coals. The PCFB combustor has performed well in terms of process characteristics such as combustion efficiency, gaseous emissions, and response to load changes. The following are the results observed for key performance parameters.

Combustion Efficiency

Testing results have shown a carbon conversion in the range of 99.8 to 100 percent with excess air levels as low as 10 percent. Very low CO levels have been observed as well.

Sulfur Retention

It has been observed that sulfur absorption in the PCFB occurs in a different manner than that in an atmospheric circulation fluidized bed (ACFB) boiler. This can result in nearly complete utilization of the sorbent. In the pilot plant testing, sulfur removal efficiencies in the range of 95 to 99.5 percent have been achieved at calcium to sulfur ratios 30 to 70 percent below what is required in an ACFB. For high sulfur coals with a Ca/S ration of less than 2, a sulfur retention of about 95 percent has been recorded.

NO_x Formation

In the pilot plant, NO_x emissions below 200 ppmvd at oxygen levels of less than 3 percent have been measured. Further reduction is possible if required by using ammonia injection. Levels below 30 ppm have been achieved with ammonia slip levels of less than 5 ppmvd.

N₂O Emissions

N₂O emissions from the pilot plant have been measured at less than 30 ppmvd at 3 percent O₂. It is expected to be lower in large size combustors where gas residence times are increased.

Hot Gas Cleanup

A key feature of the pilot plant testing has been and will continue to be the testing of ceramic barrier filter technologies. Testing was first done on an Asahi Advanced Ceramic Tube Filter and then subsequently on a Westinghouse Candle Filter. Both configurations were successful in reducing the outlet dust loading to levels required by gas turbine manufacturers, but premature ceramic element failures occurred. Evaluation of various filter modifications and additional filter designs continues.

PROJECT SCHEDULE

Budget Period 1, the preliminary engineering phase, was originally scheduled to be completed on June 30, 1993. The DOE and the project participants are currently reviewing options to reschedule the project to allow time for additional component testing. Budget Period 1 has been extended to September 30, 1993, to allow for development of these plans.

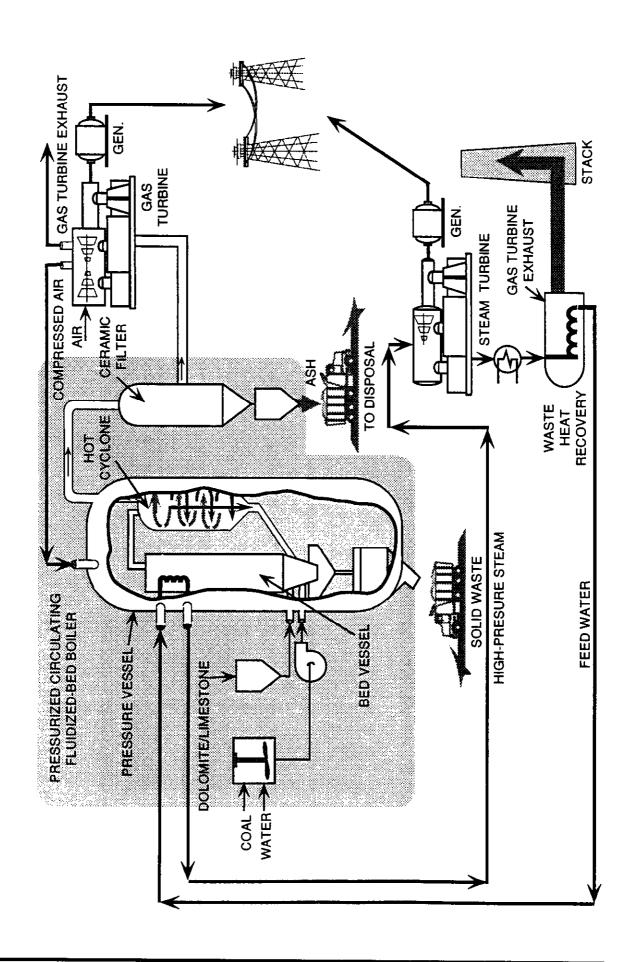
SUMMARY

The DMEC-1 Project will demonstrate the use of Pyropower's PYROFLOW pressurized circulating fluidized bed technology to repower an existing coal fired generating station. The project continues in Budget Period 1, the preliminary design phase.

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AMERICAN ELECTRIC POWER PRESSURIZED FLUIDIZED BED COMBUSTION TECHNOLOGY STATUS

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ABSTRACT

The American Electric Power Pressurized Fluidized Bed Combustion (PFBC) Program is the only ongoing PFBC and Hot Gas Clean Up (HGCU) Program in the United States. The 70 MWe Tidd PFBC Demonstration Plant is a Round 1 Clean Coal Technology Project that was constructed to demonstrate the viability of PFBC combined cycle technology. The addition of a Hot Gas Clean Up (HGCU) stream at Tidd, separately funded by the U.S. Department of Energy as an R&D project, is intended to demonstrate that Advanced Particle Filters (APF) can operate reliably in a PFBC gas stream. The experience gained from these programs is expected to hasten the commercial deployment of the technology and provide a viable power generation option in a time frame consistent with the growing baseload generation needs which are expected to develop early in the next decade.

This paper reviews PFBC technology and HGCU and discusses the status of project goals and milestones. Special emphasis is placed on the operation of the Tidd PFBC and HGCU Programs.

INTRODUCTION

The onset of the next century is expected to bring a resurgence of electric load growth. This projected increase in growth is expected to coincide with the need to replace or repower large blocks of existing capacity which will have reached the end of its useful life. The net impact will be the creation of a substantial market for Clean Coal Technologies for both new and repowering generating capacity (Figure 1.).

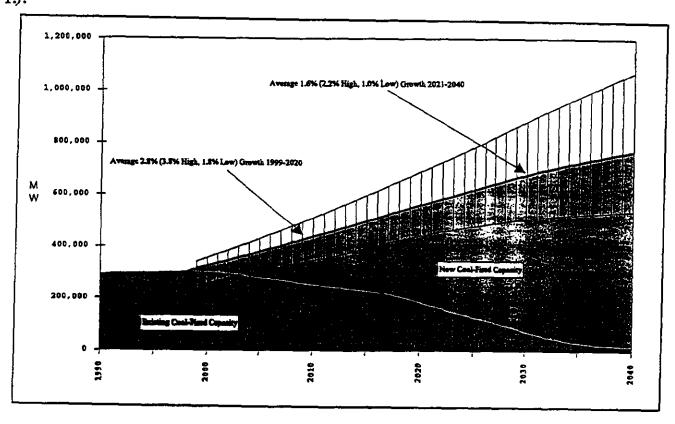


Figure 1 Potential Market for Advanced Coal-Based Technologies

This required capacity addition is expected to be met by a variety of generating options. However, coal is expected to be a dominant fuel and Clean Coal

Technologies the dominent technologies. The innovative clean coal technologies being developed and demonstrated in this decade will play an important role in meeting these power needs in an efficient and environmentally responsible manner.

TECHNOLOGY DESCRIPTION

PFBC technology is one of the advanced coal power generation options being developed. First generation PFBC technology has made significant advances over the last five years and represents an option which is ready for full scale deployment.

PFBC technology consists of a fluidized bed made up of a mass of granular particles which are maintained in a highly turbulent suspended state by an upward air flow. This fluidized state permits excellent surface contact between the air and the solid particles which permits almost isothermal conditions and efficient combustion. The temperature in the bed is established between the combustion temperature and the ash fusion temperature of the fuel – for the Tidd PFBC, this temperature is between 1520°F - 1580°F. During combustion, the SO₂ generated is removed by the addition of a sorbent such as dolomite or limestone to the bed. This process has been demonstrated to remove 90 - 95 percent of the sulfur from high sulfur coals. In addition to SO₂ removal, the process mitigates the formation of NO_x, due to its relatively low combustion temperature. The high operating pressure (approximately 175 psia) of a PFBC unit provides exhaust gases with sufficient energy to drive a gas turbine, allowing a combined cycle configuration.

TIDD PFBC DEMONSTRATION PLANT

The Tidd PFBC Demonstration Plant, a 70 MWe electric generating station in Brilliant, Ohio, is the first pressurized fluidized bed combustor to operate in combined cycle mode in the United States. Funding for the \$193-million project is being provided by Ohio Power Company, the U.S. Department of Energy (\$60.2-million) and the Ohio Coal Development Office (\$10-million).

The Tidd PFBC Demonstration involves the repowering of a 1940's vintage coal-fired power plant with PFBC components. The original Tidd Plant, consisting of two 110 MWe conventional coal-fired units, was decommissioned in 1976. The units were preserved in anticipation of a PFBC repower.

Major balance of plant equipment from the original units is utilized at Tidd. Major plant additions include the combustor building, economizer, electrostatic precipitator, and coal and sorbent storage areas.

The PFBC power island, which has been incorporated into the existing steam cycle, provides a nominal steam flow of 440,000 pounds per hour at 1300 psia and 925F, and has a gross electrical output of 70 MW. Figure 2, depicts the Tidd cycle.

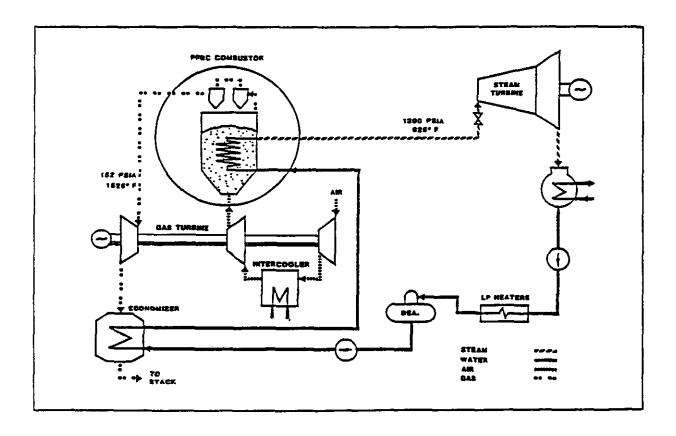


Figure 2. Tidd PFBC Demonstration Plant Cycle

Combustion air at about 175 pisa is provided by the gas turbine compressor to the combustor pressure vessel through the outer annulus of a coaxial pipe. The combustion air fluidizes and entrains bed material consisting of fuel (coal/water paste), coal ash, and sorbent (dolomite).

Seven strings of two-stage cyclones, located in the combustor vessel, remove about 98 percent of the entrained ash from the fluidized bed exhaust gases. The clean, hot gases leave the pressure vessel via the inner cavity of the coaxial pipe and are expanded through an ASEA Stal GT-35P gas turbine. The gases are exhausted through the turbine exhaust gas economizer. An electrostatic precipitator cleans the gas of particulate prior to exhausting to atmosphere.

The steam cycle is a typical Rankine Cycle with a once-through boiler. Condensate is heated in three stages of low pressure heaters and the gas turbine intercooler as it is pumped to the deaerator. A single high pressure heater and an economizer raise the final feedwater temperature to approximately 480°F. The feedwater flows through the boiler bottom zone and into the in-bed evaporator surface. Steam generated there is conveyed to a vertical separator outside the pressure vessel; flow to the separator is two-phase up to about 40 percent load and slightly superheated at full load. Saturated or slightly superheated steam from the vertical separator is routed back to the in-bed tube bundle where it passes through primary and secondary superheater sections. Final steam temperature is controlled by spray attemperation between the primary and secondary superheaters.

Coal is injected into the combustor as a coal water paste nominally containing 25 percent water by weight. Paste preparation begins by reducing the 3/4" x 0" feedstock to - 1/4" in a double roll crusher. A crushed coal recirculation system provides the ability to recirculate crushed coal to ensure correct fines content. The crushed coal is conveyed to a vibratory screen, which controls coal top size, and then into the coal water paste mixer where the appropriate amount of water is added.

The mixer discharges to two interconnected surge tanks which feed six hydraulically driven piston pumps, each of which feeds an individual in-bed full nozzle.

Sorbent feed stock sized to 3/4" x 0" is reduced to 1/8" x 0" by a hammer mill crusher. A vibratory recycle screen controls the top size of the prepared sorbent. Crushed sorbent is injected into the fluidized bed via two pneumatic feed lines supplied from dual lock hopper strings.

An alternate sorbent feed system, which provides the capability of injecting sorbent of various size directly into the coal/water paste feed system, was added in early 1993. This system provides the means to assess a wet feed sorbent system, while also providing the opportunity for better control of sorbent size consist.

Bed ash, which comprises about 50 percent of total ash generation, is removed from below the bed via a lock hopper system. Elutriated ash collected by the cyclones is removed via a pressurized pneumatic transport system which depressurizes and cools the ash without using valves or lock hoppers.

HOT GAS CLEAN UP SYSTEM

In 1992, the 10 MW (equivalent) Tidd Hot Gas Clean Up System was commissioned. This system uses ceramic candle filters to clean a portion of the exhaust gases from an operating PFBC unit. The Advanced Particle Filter (APF) is installed in a slipstream which takes one-seventh of the Tidd exhaust gases and directs these through the APF and back to the process. The HGCU slipstream replaces one of the seven secondary cyclones which is normally used for final process gas cleaning.

The HGCU slipstream is comprised of an advanced particle filter located adjacent to the PFBC combustor vessel, a back-up cyclone, a bypass cyclone, and the ancillary systems required for ash removal and ceramic filter cleaning. A schematic of the HGCU system cycle is presented in Figure 3.

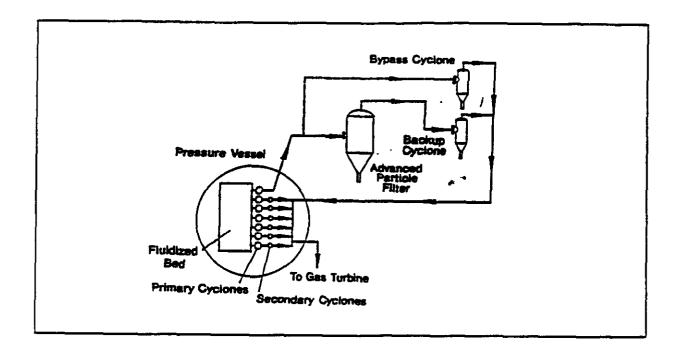


Figure 3. Tidd HGCU Test Facility Arrangement

Hot combustion gases are routed from the discharge of the primary cyclone out of the combustor vessel into the filter. The gases are then routed through the back-up cyclone and returned to the secondary cyclones' collection header located in the combustor vessel. The gases flow from this collection header to the gas turbine. A bypass cyclone is provided in the event that the APF filter is removed from service.

At full load, approximately 7600 ACFM of combustion gases at 150 psig, 1550°F, flow through the HGCU system. Normal dust loading through the filter is approximately 600 ppmw. Clean gases from the ceramic filter contain less than 15 ppmw dust loading. The back-up cyclone downstream of the filter protects the gas turbine in the event of a filter malfunction.

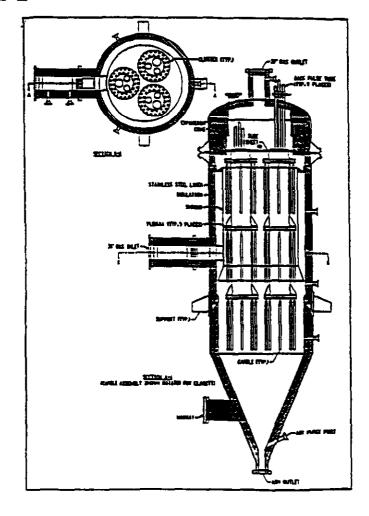
The design basis for the APF system is listed in Table 1.

AFP Design Basis				
Maximum Temperature (F)	1670			
Operating Temperature (F)	1550			
Maximum Pressure (PSIG)	185			
Operating Pressure (PSIG)	150			
Design Gas Flow Rate (LB/HR)	100,700			
Inlet Dust Loading (PPM)	500-5000			
Outlet Dust Loading (PPM)	<15			
Average Particle Size (MICRONS)	1.5			
Filter Temperature Drop (F)	5			
Pressure Drop (PSI)	3			

Table 1.

The Advanced Particle Filter, shown in Figure 4., is a 10 foot diameter by 44 foot high vessel. The vessel is internally insulated with alumina-silica ceramic insulation and lined with a 310 stainless steel liner for erosion protection. The hot gas enters at the side of the vessel and is channeled through the filter elements. Clean gas exists through the top of the vessel.

Figure 4. Tidd Advanced Particle Filter



The APF filter contains 384 ceramic candle filter elements, arranged in three clusters spaced 120 degrees apart. Each cluster holds three plenums arranged vertically, with 38 candles in each upper and middle cluster, and 52 candles in each lower cluster.

The candles are Schumacher Dia-Schumalith P40 candles consisting of a clay-bonded sintered silicon carbide support matrix coated by an aluminum silicate fibrous membrane. Each candle is 2.36 inches in outside diameter and 4.92 feet in length. The candles are attached to the tube sheet in each plenum by bolted collars and sealed by high temperature gaskets. The plenums are attached to a two inch thick, RA-333 alloy tube sheet. The tube sheet is supported from an inverted "V" expansion cone.

Candle cleaning is achieved by an air backpulse system which serves to dislodge the filter cake from the elements.

The Backpulse System is shown in Figure 5. Backpulse air is available at pressures up to 1500 psig. However, the normal backpulse pressure has been set at 800 psig.

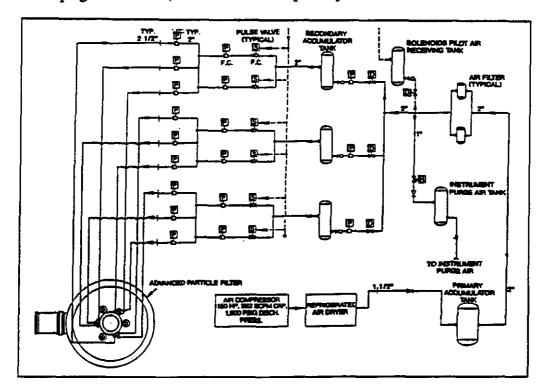


Figure 5. APF Backpulse System

Ash collected in the AFP is discharged through a pressurized screw cooler into depressurizing lock hoppers which feed a pneumatic transport system. The transport system conveys the ash from the lock hopper to the economizer outlet.

TIDD PFBC OPERATIONAL SUMMARY

The Tidd PFBC Plant achieved its first coal fire in November 1990. The details of operation through June 1992 were previously reported at the First Annual Clean Coal Technology Conference. The operating statistics for that period are summarized in Table 2.

Tidd Operating Statistics				
Through June, 1992				
Initial Combined Cycle Operation	1/29/90			
Total Operation in Coal (Hours)	2100			
Longest Continuous Run (Hours)	740			
Highest Bed Level Achieved (Inches)	142			
Highest Gross Generation (MWe)	70			

Table 2.

The Tidd Plant was removed from service in July 1992, following a successful 31-day run, for equipment repair and general maintenance. Three test runs in late July and early August 1992 ended prematurely, due to a variety of system problems. The unit was returned to service on August 8 and ran on coal for approximately 422 hours before being shutdown on August 27 for inspections. Testing focused on the feasibility of feeding sorbent with the coal-water paste.

The problems identified during initial operation of the HGCU system were generally addressed during the PFBC gas turbine outage, which began in February 1993. The system was returned to service in early July 1993. The APF system has operated for approximately 400 hours since that time. All of the problems previously identified appear to have been solved. Operation during the 400 hours run was uneventful; all of the APF system functioned as designed. Filter conditions were stable and filter pressure drop was relatively consistent. The backpulse system performed well in cleaning the candle elements of filter cake. However, difficulties are still being encountered with ash buildup on the APF hopper walls.

CONCLUSION

The Tidd PFBC Demonstration Plant has completed over 4000 hours of coal-fired operation and has generally met all performance, environmental, and reliability goals established for the demonstration. A review of the unit's operating history reveals that mechanical equipment problems accounted for the majority of system shutdowns. The first thirty months of operation have clearly demonstrated the need for a demonstration unit and have provided a clear basis for a commercial plant design. Significant strides have been made in cyclone ash removal system design and in coal preparation/coal-water paste feed system design. A clear picture is beginning to emerge with respect to system operating parameters and their impact on PFBC performance. Continued operation of the Tidd PFBC unit will continue to provide significant input to a commercial design, which will compete effectively in the repower and base load generation market.

The Tidd HGCU system has achieved almost 1000 hours of operation. Lessons learned to date emphasize the importance of auxiliary systems, especially external piping systems and ash handling systems. The design basis for such systems are being developed, applied, and refined at Tidd. Continued operation of the APF should provide an in-depth understanding of ceramic candle filter operation and

limitations. This will provide the basis for development of commercial ceramic filters capable of contributing to numerous clean coal technologies.

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Session 3 SO₂ Control Technologies

Co-Chairs:

Thomas A. Sarkus,

Pittsburgh Energy Technology Center/

U.S. Department of Energy

Lawrence Saroff,

Office of Clean Coal Technology/

U.S. Department of Energy



DEMONSTRATION OF BECHTEL'S CONFINED ZONE DISPERSION PROCESS AT PENNSYLVANIA ELECTRIC COMPANY'S SEWARD STATION PROJECT STATUS

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Abstract

The U.S. Department of Energy and Bechtel Corporation are cooperating in a joint project to demonstrate Bechtel's Confined Zone Dispersion (CZD) Technology. The demonstration is being conducted at Pennsylvania Electric Company's (Penelec's) Seward Station on Boiler Unit #15. This boiler is a 147 MWe pulverized coal-fired unit, which utilizes Pennsylvania bituminous coal (approximately 1.2 to 2.5 percent sulfur). One of the two flue gas ducts leading from the boiler was lengthened and retrofitted with the CZD technology. The new long straight duct replaced the original multi-bend duct to ensure a residence time of about 2 seconds. The goal of this demonstration is to prove the technical and economic feasibility of the CZD technology on a commercial scale. The process can achieve 50 percent sulfur dioxide (SO₂) removal at lower capital and O&M costs than other systems.

The CZD process involves injecting a finely atomized slurry of reactive lime into the flue gas ductwork of a coal-fired utility boiler. The principle of the confined zone is to form a wet zone of slurry droplets in the middle of the duct confined in an envelope of hot gas between the wet zone and the duct walls. The lime slurry reacts with part of the SO₂ in the gas and the reactive products dry to form solid particles. An electrostatic precipitator (ESP), downstream from the point of injection, captures the reactive products along with the fly ash entrained in the flue gas.

The test program is being conducted in two parts. The first part, parametric testing, started on July 10, 1991, and was completed on August 17, 1992. During this period, Bechtel's objective was to carry out a factorial test program to optimize the performance of the CZD process. The test program was designed to develop operating conditions to achieve a highly reliable and low-cost operation. The second part, from August 17, 1992 to August 30, 1993 will complete the project. Based on the results of the parametric test program, Bechtel performed additional design, procurement, installation, and facility construction as necessary to permit a 12-month continuous demonstration. The CZD system is fully instrumented and integrated with the operation of Penelec's Boiler Unit #15. The purpose of the project is to demonstrate the performance of the CZD process for SO₂ removal without significantly affecting either boiler operations or plant particulate emissions. Penelec is operating the system during the continuous demonstration as a normal part of the operation of Boiler Unit #15. Bechtel is supervising the demonstration and carrying out various performance tests, data acquisition, and chemical analyses.

Results of the demonstration indicate that the CZD process can achieve costs of \$300/ton of SO₂ removed. Based on a 500 MWe plant retrofitted with CZD for a 50 percent SO₂ removal rate, the total capital cost is estimated at less than \$30/kWe. The cost includes lime unloading, lime handling, and fully automated operation. The variable operating cost for this retrofit is estimated at less than 3.0 mills/kWh.

introduction

The CZD process involves flue gas post-treatment. It is located between the boiler outlet and the particulate collector, an ESP in most of cases.

The features that distinguish the Bechtel CZD process from other similar injection processes are the following:

- Injection of an alkaline slurry directly into the duct. Other processes use injection into a conventional spray-dryer vessel or injection of dry solids into the duct ahead of a fabric filter.
- Use of an ultrafine calcium/magnesium hydroxide, Type S pressurehydrated dolomitic lime. This commercial available product is made from plentiful, naturally occurring dolomite.
- Low residence time, made possible by the highly effective surface area of the Type S lime.
- Localized dispersion of the reagent. Slurry droplets contact only part of the gas, while the droplets are drying, to remove up to 50 percent of the SO₂. The process uses dual fluid rather than rotary atomizers.

• Improved ESP performance via gas conditioning from the increased water vapor content and lower temperatures. As a result, supplemental conditioning with SO₂ is not necessary for satisfactory removal of particulate matter.

The waste product is composed of magnesium and calcium sulfites and sulfate, with excess lime and fly ash. The waste fly ash mixture usually has pozzolanic properties. The mixture is self-stabilizing because of the excess lime and tends to retain heavy metals in insoluble forms within the fly ash.

CZD-Flue Gas Desulfurization (FGD) Demonstration as Part of DOE's Clean Coal Technology Demonstration Program

The U.S. Department of Energy and Bechtel Corporation have agreed to a cooperative effort to demonstrate the Bechtel-developed CZD technology at Pennsylvania Electric Company's Seward Station. DOE provided half, or \$5.2 million, of the project's total \$10.4 million cost. Others contributing to the project are Pennsylvania Electric Company (\$3 million), Bechtel (\$3.3 million), the Pennsylvania Energy Development Authority (\$750,000), New York State Electric and Gas Corporation (\$100,000), and Rockwell Lime Company (\$23,000). Pennsylvania Electric Company is providing the project's demonstration test site, Seward Station.

The costs of this demonstration have exceeded the \$10.4 million budget, and Bechtel Corporation has contributed an additional \$1.2 million.

Current CZD activities at Seward Station are directed toward demonstrating the best possible atomization and dispersion of the SO₂ absorbing slurry in flue gas and the performance of the existing precipitator to handle the increased dust load without adverse effects on the stack gas opacity.

The CZD project at Seward Station includes replacement of the original flue gas duct (35-foot-long segments connected with 45° elbows and corresponding turning vanes) with one new 110-foot-long straight duct ahead of the ESP.

The test program consists of two distinct periods:

- In the first period, daily factorial runs were conducted to test different atomizers, limes, and slurry concentrations. First period results were used to set and optimize the second period operations.
- In the second period, the performance of a continuously running CZD system is being demonstrated under actual power plant operating conditions. The CZD demonstration is integrated into one half of the flue gas capacity of the commercial unit (147 MW) operating continuously three shifts a day, seven days a week.

Purpose of the Test Program

The primary objectives of the project are to:

- Achieve an SO₂ removal rate of 50 percent
- Realize SO₂ removal costs below \$300/ton
- Eliminate negative effects on normal boiler operations without increasing particulate emissions and opacity

The CZD process has been automated and integrated with only one of the two existing modules of air preheater, flue gas duct, ESP, and induced draft fan associated with Unit No. 15. All auxiliary subsystems, such as lime slurrying, degritting, and lime slurry handling, have also been automated.

The demonstration project is permitting optimization of the system for application at different locations by determining the:

- Degree of atomization (slurry/compressed air ratio) versus length of duct required for evaporation of atomized slurry
- Maximum volume of slurry that can be injected per square foot of duct cross section and the confined zone dimensions of the duct cross section that will prevent deposits on duct surfaces
- Effect of flue gas inlet temperature on the evaporation characteristics, SO₂ removal, and alkali utilization

Other objectives of the demonstration project include:

- Performing comparison tests of hydrated calcitic lime and freshly slaked calcitic lime
- Testing methods for improving ESP performance during lime injection by:
 - Monitoring ESP operating and opacity variations during all injection tests
 - Performing particulate emission measurements on several extended runs
- Testing different slurry atomizers to determine the most energy-efficient and erosion-resistant
- Testing the effect of burning higher sulfur coal on the percentage of SO₂/NO_x removal

Description of the CZD Process

The spray of lime slurry is injected close to the center of the flue gas duct parallel to the flow of gas. As a cone of spray moves downstream and expands, the gas within the cone cools and its SO₂ is rapidly absorbed by the liquid droplets.

Spray droplets on the outside of the cone mix with the hot gas and dry very rapidly. With the proper choice of slurry concentration and injection rate, drying will be complete before the droplets contact the walls of the duct. The process does require a sufficient length of straight duct downstream from the sprays, estimated at 100 feet, and the gas flow must be reasonably uniform where the spray is injected. Judicious use of the turning vanes, typically installed to minimize pressure drops, makes the gas flow in the bends more uniform.

By carefully positioning lime slurry atomizers, it is possible to obtain a wet zone in the middle of the duct with an envelope of hot gas between the wet zone and the duct walls. This is the principle of the confined zone.

Gas velocity in large ducts is generally about 60 fps at full load, and the flow is highly turbulent. Thus, spray droplets in the expanding cone are transported outward by eddy diffusion. However, since the outward cone's diffusing droplets continuously contact hot gas at about 300°F, they rapidly achieve surface dryness. Exposed to the highly localized full concentration of SO₂, the lime reacts extremely rapidly.

At a certain point downstream, the free moisture in the spray will evaporate completely, and the remaining dry solids can contact surfaces of the duct or the turning vanes without adhering and causing deposits to accumulate.

For removal of 50 percent of the SO₂ from flue gas with a slurry of pressurized hydrated dolomitic lime (PHDL), slurry concentration is a major variable. Enough slurry must be added to achieve the desired results. The demonstration program provides an opportunity to explore and optimize the control of this variable.

The chemical mechanism required for the absorption of SO₂ from the flue gas is simple and very well known. In the presence of water, SO₂ from the flue gas is absorbed as sulfurous acid:

$$SO_2 + H_2O \rightarrow H_2SO_3$$

In the presence of water, pressure-hydrated dolomitic lime reacts instantaneously with H₂SO₃, producing calcium and magnesium sulfites and sulfates:

$$Ca(OH)_2$$
 . $Mg(OH)_2$ + $2H_2SO_3$ \rightarrow $CaSO_3$ + $MgSO_3$ + $4H_2O$
 $CaSO_3$ + $1/2O_2$ \rightarrow $CaSO_4$
 $MgSO_3$ + $1/2O_2$ \rightarrow $MgSO_4$

Past CZD Experience

Over the last few years, considerable testing of the CZD technology was performed as proof-of-concept on pilot and commercial units. The encouraging results led to the

new phase of demonstrating the process in a commercial unit. References 2, 3, 4, and 5 describe the test programs and the test results of the earlier work.

Overall System Description

Figure 1 shows a simplified, overall flow diagram of the Seward CZD system and of the Boiler Unit No. 15 flue gas system. The two systems are closely linked. The boiler has twin air and flue gas systems, designated "A" and "B." The CZD system removes SO₂ from the B flue gas stream.

The overall CZD system includes the following process operations and supporting functions:

- Lime slurry preparation
- Lime slurry feed
- Lime slurry injection
- FGD duct (flue gas desulfurization section)
- Atomizing air compression

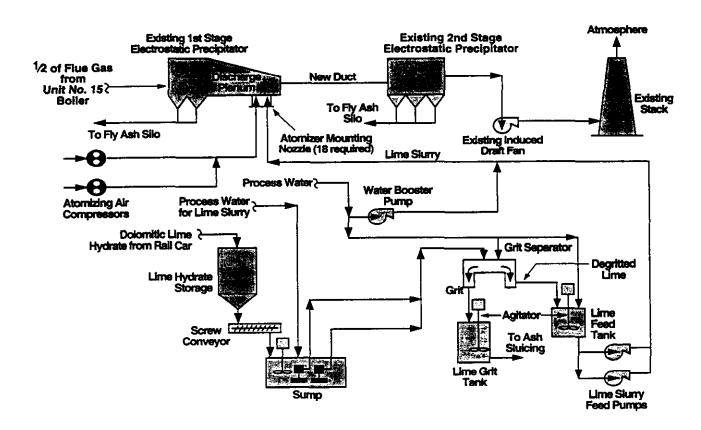


Figure 1 Seward Station Overall Process Flow Diagram

Figure 2 depicts the interrelationship between individual process operations and supporting functions.

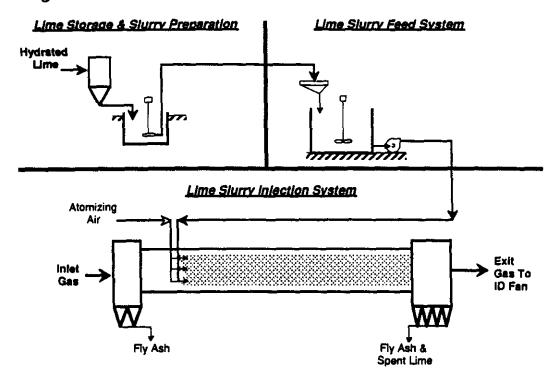


Figure 2 Seward CZD System Overview

Lime Slurry Preparation

The lime slurry preparation system contains:

- A 50-ton lime silo for receiving and storing lime hydrate, with a vent baghouse filter
- A 5,000-gallon lime hydrate slurrying sump with an agitator
- A rotary air lock valve driven by a variable-speed motor and a screw conveyor for transferring the lime hydrate from the lime silo to the sump
- Two sump pumps, one working and one spare, for transferring the lime slurry to the CZD feed system

Receiving and Storing Lime Hydrate

The lime silo has enough capacity for one day of lime usage. Consequently, daily deliveries of lime are necessary. This silo was recently upgraded for use in the CZD system. Its vent baghouse filter was fitted with new bags and its high- and low-level probes were provided with high- and low-level alarms. The rotary air discharge valve for this silo was equipped with a variable-speed motor for controlling the discharge on the hydrate to the lime slurrying sump. The speed of this rotary valve is controlled by the slurry sump density controller.

Slurrying of Lime Hydrate in the Lime Sump

The lime slurrying system was designed for fully automatic operation governed by the level controller in the lime feed tank. One of the two sump pumps operate continuously, pumping the lime slurry to the CZD lime feed system. The slurry level in the tank governs the transfer of lime slurry from the sump to the lime feed tank. The tank level controller tends to maintain a constant level in this tank by the operation of a lime flow control valve in the lime transfer line from the sump to the vibrating screen. As the transfer of the lime slurry varies, the lime slurry level in the lime sump also varies.

The lime sump is equipped with a level controller designed to maintain a constant level of slurry in the sump by controlling the sump's water inflow.

The lime sump pump bypass is equipped with a lime slurry density controller which maintains a constant concentration of lime slurry in the sump by controlling the discharge rate of lime hydrate from the silo (speed of rotation of the air lock discharge valve).

Figure 3 shows the lime storage and slurry preparation required for continuous operation.

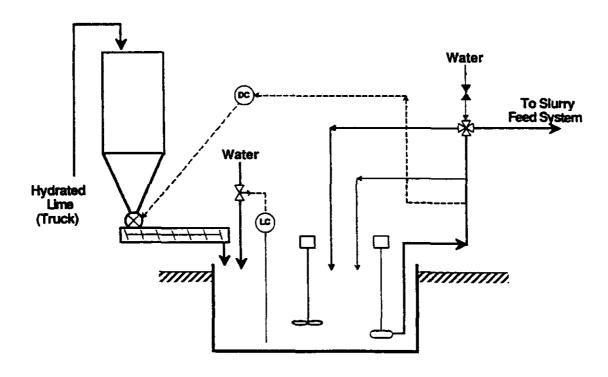


Figure 3 Lime Storage and Slurry Preparation - Continuous Operation

Lime Slurry Feed

The lime slurry feed system consists of:

- One vibrating screen for the removal of oversized materials from the lime slurry
- Two grits slurry tanks, one working and one spare, both equipped with agitators and level indicators
- Two lime slurry feed tanks, one working and one spare, both equipped with agitators, level controllers, and temperature indicators
- · Two lime slurry feed pumps, one working and one spare

The system is designed for intermittent parametric testing as well as continuous plant operation; hence it has double tankage.

The vibrating screen is designed to degrit the lime slurry and is used for the removal of foreign matter from this slurry (sand, trash, etc.). Foreign material drops from the vibrating screen into the collecting gutter from which it is sluiced with water into the grits tank.

The filtered lime slurry is discharged from the vibrating screen into the lime slurry feed tank. The slurry level in the tank is controlled by the tank level controller, which throttles the flow of lime slurry from the lime slurry sump pump to the vibrating screen. The lime slurry feed pump is used to pump the lime from the feed tank to the lime slurry injection header. Figure 4 shows the lime slurry feed system – continuous operation.

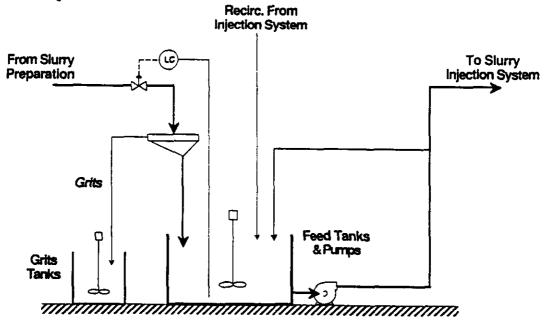


Figure 4 Lime Slurry Feed System - Continuous Operation

Lime Slurry Injection

The lime slurry injection system consists of:

- The lime slurry and water piping
- The flow controls on the top of the desulfurization duct (other than the lime slurry distribution header and atomizer feeders)
- The water booster pump and associated water piping at ground level

Lime slurry is supplied to the injection lime header from the lime feed system via the loop main, which consists of the feed supply and the excess feed return headers. Operation of the atomizers requires relatively high, constant, lime slurry injection pressure. This pressure is maintained at a constant level at the inlet to the injection header by the back pressure controller in the lime slurry return header. The flow of the lime slurry to the atomizers' distribution header is controlled by a flow controller which is reset by the Section C temperature controller. In the duct, Section C is in the turning vanes area.

The lime slurry injection header is connected to the lime slurry feed loop via a four-way valve. This valve connects the lime injection header to the water supply piping from the water booster pump. The use of the four-way valve permits the lime injection header to be flushed with water whenever the lime injection is interrupted. The water supply header is furnished with a water flow controller that can be reset by the Section C temperature controller. Both the lime and water flow controller valves are connected to the low-pressure switch on the atomizing air supply header so that the operating flow control valve(s) will close in case of low atomizing air pressure. The arrangement protects the flue gas handling system from being flooded with unatomized lime slurry or water.

The power plant domestic water distribution system provides water for flushing the atomizers and their lime slurry supply piping and for injection into the flue gas stream. Because the operating pressure of this system is inadequate for the operation of atomizers, the CZD injection system is equipped with a water booster pump to maintain an adequate water supply pressure.

Figure 5 shows the lime slurry injection, continuous operation.

Flue Gas Desulfurization Section

Seward Boiler Unit No. 15 is a balanced draft boiler provided with two F.D. fans, two Ljungstrom air heaters, two twin-chamber ESPs, and two I.D. fans. The two ESPs are joined by twin flue gas ducts that form twin flue gas treating trains, referred to as A and B trains.

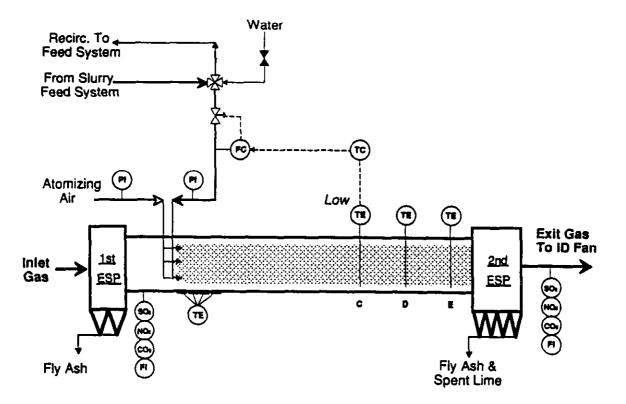


Figure 5 Lime Slurry Injection - Continuous Operation

The desulfurization duct has a 110-foot-long straight run for injection of the atomized lime slurry. This duct length is necessary for the boiler unit for the absorption of SO₂ from flue gas and for drying out the absorption products. The atomizing nozzles are located at the duct inlet.

The operating instrumentation includes a low-pressure switch which will stop lime or water from being injected if the air pressure is too low to ensure adequate atomization. This instrumentation is essential for the protection of the flue gas system from the formation of wet deposits, plugging, and flooding.

A ready/standby switching system allows the lime slurry feed to the atomizers to be diverted back to the feed tanks, while water is supplied to flush the atomizing nozzles and lime supply header. The ready/standby system can be used to temporarily suspend lime injection without shutting down the CZD system and can be activated from the plant control room.

Atomizing Air Compression System

This system contains two screw-type air compressors (which can be operated singly or in parallel) and an air receiver. Each of the two compressors can supply up to 2,250 scfm of air at 120 psig and is driven by a 500 bhp motor. Each compressor is equipped with air intercoolers and after-coolers using 100 gpm of cooling water. The

compressors are of the oilless type and provide oil-free compressed air. The operation of the CZD system requires continuous operation of at least one of the two compressors.

Instrumentation and Control (I&C) System

Instrumentation and control (I&C) is broken down according to the plant locations at which CZD equipment and systems are found. These five operational areas/systems are:

- Lime slurry preparation system
- Lime slurry feed system
- Lime injection system
- Duct flue gas monitoring and controls
- Atomizing air compression system

Startup, operation, and monitoring of the equipment and systems within these areas are accomplished by a combination of actions performed locally or in the control room (remote operation). In general, initial startup of all pumps, mixers, and systems must occur locally. In this way the operator can visually verify the condition of the equipment in the area and determine whether it is safe to put the equipment or system into operation. Once a system or equipment is in operation, monitoring the condition of equipment and changing the system setpoints can be done remotely in the control room, or locally through panel-mounted switches and controllers.

The ready/standby system is also part of the CZD I&C and operates through the Bailey Distributed Control System (DCS). The ready/standby switch gives the operator a means of controlling whether or not lime slurry is injected into the duct without unnecessarily upsetting CZD controls, and safeguards the operation of the Buell ESP. Low atomizing air pressure also activates the standby mode of operation.

Remote monitoring and control of the CZD process from the control room are provided by the existing combustion management control system (MCS) and are supplemented by the process control view station (PCV). Additional plant and process operating information is available from Leeds & Northrup (L&N) recorders located in the ESP control room and in the duct B instrumentation room.

Parametric Test Results

The parametric tests included duct injecting atomized lime slurry made of dry hydrated calcitic lime, fresh slaked calcitic lime, and pressure-hydrated dolomitic lime. All three reagents removed SO₂ from the flue gas, requiring different concentrations in the lime slurry for the same percentage of SO₂ removal. The most efficient and easiest to operate is the pressure-hydrated dolomitic lime. The lime

slurry duct injection does not adversely impact the stack opacity. On the contrary, it substantially reduces stack opacity during the lime injection.

Continuous Operation Results

Table 1 shows typical results when using pressure-hydrated dolomitic lime.

Table 1 Typical Results
Continuous Demonstration Tests with Pressure-Hydrated Dolomitic Lime (PHDL)

Date	8/21/92	8/21/92	8/24/92	8/26/92	8/26/92	8/27/92	11/12/92
Item Time	21:15	22:15	18:15	14:45	21:15	5:45	15:30
Boiler Load, MW	139.5	144.5	143.5	144.4	142.8	142.6	120.4
Stack Opacity, percent	7.8	7.8	11.5	6.8	10.2	5.9	14.2
Flue Gas Temp. in, °F	305.2	304.4	309.8	310.2	304.2	305.4	284.7
Flue Gas Temp. out, °F	192.1	190.9	191.1	197.8	190.7	194.1	197.3
SO ₂ in, ppm	910	910	869	905	944	901	<i>7</i> 03
SO ₂ out, ppm	333	351	343	401	368	399	377
Lime Slurry Flow Rate, gpm	55.8	54.9	58.9	52.4	55.3	54	37.7
Lime Slurry Conc., percent	9.0	9.0	12.0	6.4	12	6.5	10
SO ₂ Removed, percent	55.3	53.1	50	48	51.2	45.3	39.3
Lime Utilization, percent	41.0	40.1	23.7	49.8	27.8	46.4	25.3

It is clear that a high lime slurry injection rate is required to achieve a high percentage of SO₂ removal. Following are the conditions that limit the injection rate:

- The flue gas flow rate and temperature
- The residence time in the straight duct

The design of the CZD system at Seward Station is based on the following assumptions:

- An inlet flue gas temperature in the B-duct of 300°F
- A boiler load of 147 MW, equivalent to a flue gas velocity in the duct of 55-66 ft/sec
- A 110-foot-long straight duct, equivalent to 2 seconds' residence time

During the second half of August 1992, the flue gas temperature was 300-310°F, the boiler load was 145-147 MW, the residence time in the duct was 2 seconds, and the injection of lime slurry was 52-57 gpm with SO₂ removal above 50 percent.

Figure 6 shows the CZD duct temperature versus the lime slurry injection rate and percentage of SO₂ removal during the period August through November 1992. In order to include the duct inlet temperature in this graph, it has been divided by 5—i.e., 300°F is represented by 60°F. The graph shows that during the months of October and November, the inlet flue gas temperature continued to descend from above 300°F to 260-280°F due to air leakage in the boiler air heater. This low temperature limited the lime slurry injection rate to 30-40 gpm and resulted in a 15-30 percent variation in the percentage of SO₂ removal.

The percentage of SO₂ removal is dependent on the lime slurry injection rate and that the percentage of lime slurry concentration above a specific level does not affect the percentage of SO₂ removal.

Figure 7 shows SO₂ removal versus the lime slurry injection rate, expressed in gpm and based on daily averages during the period August 17 to September 16, 1992. This graph shows that for a low lime slurry injection rate — for example, 36 gpm — the percent of SO₂ removal was only about 20 percent. By increasing the injection rate to 50 gpm, SO₂ removal increased to 38 percent. At a 54 gpm injection rate, the SO₂ removal rate increased to 45 percent. By extrapolation, with a 60 gpm injection rate, the percentage of SO₂ removed will reach 55 percent. As mentioned above, the percentage of SO₂ removed represents daily averages and not peak values.

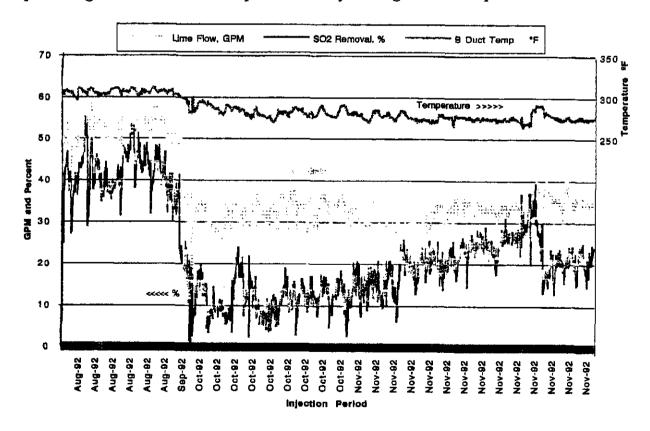


Figure 6 Temperature vs Lime Flow and SO₂ Removal August 17 to November 16, 1992

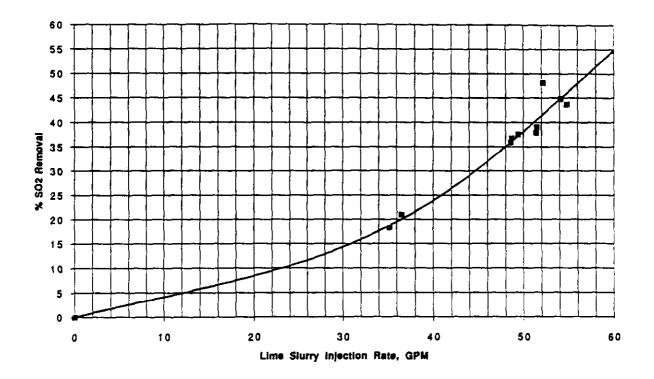


Figure 7 Removal vs Injection Rate, Based on August 17 to September 16, 1992 Daily Rates

Chemical Analyses of Fly Ash and CZD Reacted Products

Samples of solids containing fly ash, CZD reacted products, and unreacted lime were collected and analyzed. The samples were collected during the CZD continuous operation with pressure-hydrated dolomitic lime. The analyses were made by scanning electron micrographs (SEM) and energy-dispersive X-ray analyses (EDAX) for elements in fly ash and CZD reacted products. The analyses were performed at Pennsylvania State University, Department of Material Science and Engineering, Fuel Science Program.

Figure 8 is an SEM and Figure 9 is an EDAX of a sample of fly ash with CZD reaction products. The fly ash particles on the SEM appear as gray spheres 1-20 microns in diameter. The reaction products (calcium sulfite, calcium sulfate, and magnesium sulfate) appear as separate white crystals from submicron size to about 2 microns in diameter. The very fine crystals of CZD reaction products agglomerated on the larger fly ash spheres, creating larger particles that are more easily removed by the ESP from the flue gas.

The EDAX analyses show the main elements and their concentration. Position 5, shown in Figure 9, shows calcium, magnesium, sulfur and oxygen as principal elements of the reaction products. Other positions show silica, aluminum, iron, and titanium as main elements (the fly ash constituents), and calcium, magnesium, and sulfur (reaction-product constituents) at lower concentrations.

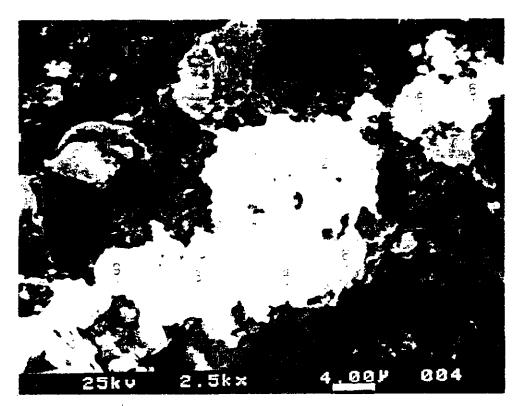


Figure 8 Scanning Electron Micrograph of Samples of Fly Ash, CZD Reaction Products, and Unreacted Lime

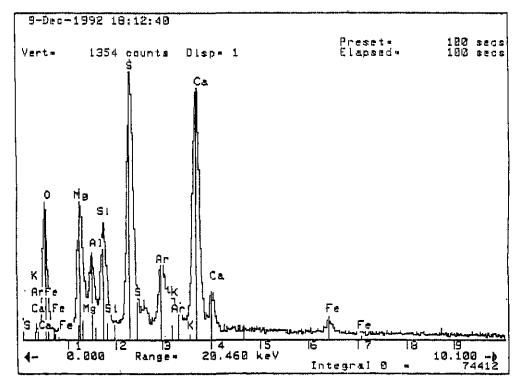


Figure 9 Energy-Dispersive X-ray Analysis of Position 5 in Figure 8

Percentage of Lime Utilization and the Factors that Could Affect It

The percentage of lime utilization in the CZD system significantly affects the total cost per ton of SO₂ removed. For this reason, we have examined methods that can improve the percentage of lime utilization. An analysis of the continuous operational data compiled to date indicates that the percentage of lime utilization is directly dependent on two factors:

- The percentage of SO₂ removal
- The lime slurry concentration

For operating conditions at Seward Station, data show that a 40-50 percent SO₂ removal and a 6-8 percent lime or dolomitic lime slurry concentration will assure a 40-50 percent lime utilization rate. That is, 2 to 2.5 moles of CaO or CaO.MgO are required for every mole of SO₂ removed; or assuming 92 percent lime purity, 1.9-2.4 tons of lime are required for every ton of SO₂ removed.

Technology Applicability and Limitations

Commercial Application

CZD technology is particularly well suited for retrofitting existing boilers, regardless of type, age, and size, type of coal burned, or the percentage of sulfur in the coal. Compared to currently available flue gas desulfurization systems, CZD technology can be more easily and economically integrated into existing power plants.

The inherent advantages of the CZD process relative to currently available commercial technologies are:

- Substantially lower capital cost and total cost per ton of SO₂ removed
- Ease of retrofit because CZD eliminates the need for chimney alterations, boiler reinforcements, and modifications to boiler draft controls
- No increase in flue gas pressure drop; therefore, no extra fans are needed
- Minimal space requirements in the stack area
- No dewatering or liquid waste treatment required
- No flue gas reheating requirement
- No congestion close to the boiler or stack
- Easily disposable reaction products that are dry, free flowing, and are removed with the fly ash
- · Reduced labor and maintenance requirements
- Applicable to wide range of: geographic locations, load profiles, and particulate collectors (either ESP or baghouse)

 Raw materials (dolomitic rock is well distributed in many areas of the United States)

A disadvantage of the CZD process is the limitation in SO_2 removal to about 50 percent.

The CZD process requires that drying and SO₂ absorption take place within 2 seconds. A long straight flue gas duct of about 100 feet is required to ensure a residence time of 2 seconds. If this length of flue gas duct is not available and there are space limitations for a new horizontal long duct, then a vertical duct of the required length can be built to enable utilization of the CZD technology.

Reference 6 describes flue gas desulfurization by the CZD process on a comparative basis with the economies of other clean coal technologies.

Commercial Demonstration

The CZD project is designed to demonstrate:

- Reliable operation of the CZD process when integrated with a power station
- Absence of any detrimental effect on normal boiler operations
- Capability to operate with high- and low-sulfur coal

Bechtel intends to commercialize the CZD process when the present demonstration is successfully concluded. During the CZD demonstration, papers giving technical and economic data, results, and conclusions, will be presented at various conferences. These papers will be made available for publication in appropriate journals of technical societies, the electric generating industry, and in other publications. Representatives of utilities will be invited to visit the demonstration site and learn how SO₂ can be removed cost-effectively using the CZD process.

Conclusions

- Parametric tests with key variables were clearly determined and used as a basis for the continuous operation system
- The fully automated and integrated system with the power plant operation proved that the CZD process responds very well to automated control operations
- CZD system availability is very good. For example, during the period October 17 to November 16, 1992, the automated systems were on line 100% of time – 744 hours

- During normal CZD operations, no deposits of fly ash and reaction products took place in the flue gas duct. On a few occasions, deposits accumulated in the duct and had to be manually removed. The failures that caused these deposits were identified as follows:
 - A power failure on the air compressor supplying the compressed air to the lime slurry atomizers
 - Breakage of the ceramic tips, causing coarse sprays
 - Leakage of the lime slurry atomizers around their flanges and spray tips, causing the formation of deposits on the atomizer tips

These problems were corrected, and we expect that they will not be repeated. A new control system will monitor the operation of each atomizer and will stop operations if one of them is not operating correctly.

- At Seward Station, stack opacity is not detrimentally affected by the CZD system
- Results of the demonstration indicate that the CZD process can achieve costs of \$300/ton of SO₂ removed when operating a 500 MW unit burning 4% sulfur coal. Based on a 500 MWe plant retrofitted with CZD for a 50 percent rate of SO₂ removal, the total capital cost is estimated at less than \$30/KWe

 the first-of-a-kind agreement (Own-and-Operate) to design, engineer, construct, fabricate, own, operate, maintain and finance a power plant scrubber, by a nonutility.

These and other features allow the scrubber to have improved environmental performance, reduced space requirements, better energy efficiency, and lower costs than conventional first (or second) generation scrubbers. With specific regard to environmental management, this project seeks to demonstrate that air pollution control need not have deleterious solid waste and/or wastewater consequences.

Construction of the scrubber is complete; operations began in June 1992, ahead of schedule and within budget. The Clean Coal demonstration project calls for three years of operations. After the three-year demonstration period, Pure Air will continue to Ownand-Operate the scrubber for the next 17 years.

This paper reviews the advanced wet flue gas desulfurization (FGD) design features, and the environmental and business features of the project. This paper also includes operations data, project costs and schedule.

Background

Pure Air began development efforts in early 1988 for an Own-and-Operate Advanced FGD facility serving NIPSCO. With the cooperation of NIPSCO, the project was submitted to the United States Department of Energy for consideration under the Innovative Clean Coal Technology Program (Solicitation II).

Pure Air's "Advanced Flue Gas Desulfurization (AFGD) Demonstration Project" was one of 16 projects selected under DOE's second Clean Coal solicitation. Following selection in September 1988, DOE entered into negotiations with Pure Air, and a Cooperative Agreement was awarded on 20 December 1990.

In September 1989, a flue gas processing agreement was signed with NIPSCO, whereby an AFGD facility would be constructed at its Bailly Generating Station located in Porter, Indiana, forty miles east of Chicago on the southern shore of Lake Michigan. The project has since progressed through design and construction, and into operation. Indiana law required that NIPSCO obtain from the Indiana Utility Regulatory Commission (IURC) a "Certificate of Public Convenience and Necessity". This law was originally written to create incentives for utilities to install clean coal technology. The Certificate process was initiated on 15 August 1989 and a certificate was granted on 11 April 1990. In doing so, the IURC concurred that by installing the AFGD facility at Bailly, NIPSCO was using the most cost-effective solution for SO2 reduction.

Bailly Station consists of two coal-fired boilers — Unit #7, rated at 183 MW net capacity, went into operation in 1962; and Unit #8, rated at 345 MW net capacity, started in 1968. Both units burn bituminous coal, typically ranging from 2 to 4 percent sulfur.

First, a brief description of the plant and the installed FGD system (i.e., absorber design, limestone feed system, gypsum dewatering and wastewater treatment system) are outlined. Next, a summary of the results of the performance testing completed on the AFGD system is presented. Also included is the DOE demonstration test matrix and results of the first DOE test conducted in September 1992, along with relevant operations information.

Plant Description

A simplified process flow diagram for the Bailly Advanced FGD plant is illustrated in Figure 1.

This facility includes a single absorber designed to process the maximum quantity of flue gas generated from both NIPSCO's Units #7 and 8. The absorber is equipped with a co-current open grid tower with two levels of slurry distribution, an integrated reaction tank and a two-stage vertical mist eliminator in the horizontal duct. The absorber is designed to accomplish several process steps, including gas quenching, absorption of SO2, reaction with limestone and oxidation of sulfite to gypsum in a single vessel. The co-current absorber is designed at higher than conventional counter-current flue gas flow rates which makes the co-current vessel very compact. The flue gas from both boilers (approximately 2,200,000 scfm) through the existing I.D. fans is combined in a common duct feeding a single absorber tower where it contacts the recirculation slurry. Quenching and humidification of the gas and absorption of the SO2 occur simultaneously. The zone where the flue gas initially contacts the recirculation slurry is called the "wet-dry" interface and is washed intermittently with fresh water to prevent the formation and growth of deposits.

Recirculating slurry is split between the two levels of distribution provided. The slurry and flue gas pass co-currently over the open grid packing located in the absorber tower. The grid packing, made of polypropylene, provides a large surface area for gas-to-liquid contact to enhance the SO2 removal efficiency. The SO2 of the flue gas is absorbed into the slurry, and the amount of gas phase SO2 reaction is reduced as the gas flows through the tower. Partial oxidation of the absorbed SO2 occurs in the tower. Oxidation of the SO2 is completed in the reaction tank. After flowing downward through the absorber tower, the scrubbed flue gas makes a 90° turn and passes over the liquid in the reaction tank where gas-liquid disengagement occurs. The gas exits the absorber by passing through a highly efficient, two-stage mist eliminator located vertically in the horizontal outlet duct, where the cleaned flue gas exits through a newly built stack. The reaction tank is located beneath the absorber tower, so the recirculating slurry with absorbed SO2

falls directly into the tank. The reaction tank is designed to hold an adequate liquid volume to ensure efficient usage of limestone and to provide adequate residence time for complete oxidation of calcium sulfite to calcium sulfate.

Three air rotary spargers (ARSs) are provided to maximize the efficiency of the oxidation of calcium sulfite. The patented ARS is an innovative and unique design which combines the process functions of agitation and oxidization. By simultaneous rotation of the ARS and the introduction of air through the sparging nozzles, very fine bubbles are produced.

In addition to the three ARSs, a small section of the absorber reaction tank is also fitted with two rows of fixed air spargers (FAS). Incursion of the FAS helps ensure complete oxidation at maximum levels of coal sulfur.

To neutralize the absorber slurry, dry pulverized limestone is pneumatically conveyed to the absorber reaction tank. The SO2 content in the flue gas at the stack is monitored and controlled by regulating the quantity of limestone injection into the reaction tank.

During humidification of the flue gas, water is consumed from the reaction tank by evaporation. To ensure that the solids concentration in the reaction tank is maintained at 25 percent, slurry is transferred from the reaction tank to the gypsum dewatering equipment.

Summary of Project Features

Large, Single-Module Scrubber

The AFGD facility is sized to scrub all of the flue gases from both of the Bailly Station's two coal-fired units. Unit 7 has a nameplate capacity of 194 MWe, and is permitted at 183 MWe. Unit 8 has a nameplate capacity of 422 MWe and is permitted at 345 MWe. Thus, the AFGD facility is currently operating at 528 MWe (or 1,420,000 SCFM); but it is sized to handle Bailly's nameplate capacity of 616 MWe, if the power plant is repermitted. This point is important to keep in mind, when performing economic assessments.

With a nominal capacity of 600 MWe, the Bailly scrubber is the largest single SO2 absorber in the United States. Under the Clean Air Amendments of 1970, as amended by the Clean Air Act Amendments of 1977, newly constructed power plants were required to scrub SO2 emissions; otherwise, they would not be allowed to operate. As a result, the conventional U.S. practice during the 1970s and early 1980s was to install several small absorber modules, including one or two spare/backup modules. (For example, four 175 MWe modules might be installed at a 600 MWe power plant.)

The AFGD facility at Bailly Station is the first large (i.e.,>500 MWe) single-module absorber in the United States, and the largest in the world to operate with high-sulfur coal. It has no spare or backup modules, and treats all of the combined flue gases from Units 7 and 8 at the Bailly generating station.

High Velocity, Co-Current SO2 Absorber

As the elimination of spare/backup modules serves to reduce the scrubber's cost and relative size, so does its advanced design. Figure 1 is a simplified flow diagram of the AFGD system at Bailly Station. Pure Air utilizes a co-current absorber, in which the limestone slurry moves in the same direction as the power plant flue gases. This enables the Bailly scrubber to operate at a flue gas velocity of approximately 18 to 20 feet/second, versus 8 to 10 feet/second for a more conventional counter-current scrubber. As a result of the higher flue gas velocity, the absorber vessel is smaller, with commensurate cost savings.

Direct Limestone Injection

At the Bailly scrubber, pulverized limestone is injected directly into the SO2 absorber. This direct injection of pulverized limestone eliminates the need for on-site wet grinding systems, thereby reducing both space requirements and capital costs. Direct limestone injection is particularly attractive at power plants with limited space availability. The Bailly scrubber is one of only a few scrubbers in the world to utilize direct limestone injection, and the first in the United States. The direct limestone injection system has operated without problems to date.

High Quality Gypsum By-Product

Conventional first and second generation scrubbers produced a mixture of calcium sulfite (CaSO3) and calcium sulfate (gypsum, CaSO4). This mixture is commonly called "scrubber sludge". It caused scaling problems in many early scrubbers, and must be stabilized prior to its disposal as a solid waste.

Most of today's scrubbers either inhibit the oxidation of CaSO3 or force the oxidation to gypsum, because scaling problems are generally not encountered with either pure CaSO3 or pure gypsum. Forced oxidation to gypsum has a potential advantage over inhibited oxidation in that the gypsum by-product can often be utilized commercially, depending upon market conditions.

For the AFGD project, NIPSCO has entered into a long-term contract with U.S. Gypsum (USG), whereby USG is purchasing all of the by-product gypsum for use as a feedstock at its East Chicago wallboard manufacturing plant. Wallboard-grade gypsum specifications are shown in Table 1, along with the Bailly gypsum characteristics measured during a 100-hour performance test in August 1992. From start-up through the end of June 1993, the AFGD project produced 216,344 tons of wallboard-grade gypsum.

Note that scrubber gypsum can be used in other applications such as cement, and the gypsum purity requirement is generally lower for cement than for wallboard.

Wastewater Treatment and Evaporation Systems

The Bailly generating station is situated on the southern shore of Lake Michigan, and the AFGD facility utilizes process water taken from the lake. Much, though not all, of the process water is recycled within the AFGD system. Treated wastewater is discharged into Lake Michigan. Wastewater requirements are shown in Table 2, along with measured wastewater characteristics at the Bailly Station.

Chloride content is a critical parameter for wallboard-grade gypsum. Removal of chlorides from the gypsum can be accomplished easily by washing the gypsum, but the resultant wastewater can often exceed permit requirements. To avoid this potential problem, Pure Air will demonstrate a novel Wastewater Evaporation System (WES). Part of the process water stream is bled off to maintain an acceptable chloride level within the absorber vessel, and then injected into the power plant ductwork where hot flue gases evaporate the water. Upon evaporation of the water, any dissolved salts will crystallize, so that they can be collected along with fly ash by the power plant's particulate control devices. The salts are then easily disposed of with the power plant fly ash.

At the Bailly Station, the WES will be demonstrated only on Unit 8 (422 MWe nameplate; 345 MWe permit capacity). Taken together, gypsum utilization and wastewater evaporation will demonstrate that SO2 control need not have deleterious solid waste and/or wastewater consequences.

PowerChipTM Gypsum Demonstration

The AFGD by-product gypsum is in a finely powdered form. However, Pure Air will demonstrate a process to agglomerate and flake part of the by-product gypsum stream, in an attempt to improve the marketability of scrubber gypsum to end-users who are accustomed to using natural gypsum rock. This PowerChip gypsum can be transported more easily and handled with existing equipment at most wallboard and/or cement plants. As an add-on to the AFGD project, Pure Air will attempt to blend fly ash and wastewater treatment solids into the PowerChip gypsum by-product. Although these impurities would make the gypsum unacceptable for wallboard applications, it could still be used in cement. Pilot tests have indicated that maximum fly ash loadings of 20% to 30% may be achieved. In combination with wastewater evaporation and the coproduction of wallboard-grade gypsum, this process may bring coal-fired power generation technology one step closer to the goal of a zero-discharge power plant. Demonstration of the PowerChip gypsum process is scheduled to begin in mid-1993.

Own-and-Operate Business Arrangement

The AFGD project marks the first time that a power plant scrubber is owned and operated by someone other than the host utility company. Under this so-called "own-and-operate" business arrangement, Pure Air is responsible for financing, designing, constructing, operating and maintaining the scrubber. NIPSCO pays a monthly service charge to Pure Air. This arrangement allows NIPSCO to focus on its core business of electricity generation and distribution, while Pure Air specializes in scrubber design, construction and operation. Under the terms of a flue gas processing agreement between Pure Air and NIPSCO, Pure Air will operate the AFGD system for an additional 17 years, after the three-year demonstration project is completed.

Summary of Project Operations

To date, operations have gone well. The scrubber has already exceeded its target of demonstrating 95+% SO2 removal capability, while producing a commercial gypsum by-product. From start-up 2 June 1992 to 15 June 1993, the AFGD facility removed 76,540 tons of SO2 at the Bailly Station. Current operations are largely uneventful. Some key operating data are shown in Table 3. Future operations will be punctuated by a series of demonstration tests.

Project Costs

The budget and costs for the AFGD project are summarized in Table 4. The total project budget, including the PowerChipTM gypsum demonstration, is \$151,707,898. Of this amount, DOE is funding \$63,913,200, or 42%. Design and construction of the nominal 600 MWe AFGD facility were completed slightly under budget, prior to the addition of the two-year PowerChip gypsum demonstration, which will cost about \$1.2 million.

Project Schedule

Groundbreaking for the AFGD facility was held on 20 April 1990, which coincided with the twentieth anniversary of Earth Day. On 2 July 1991, a major accident occurred at the project site when two 14 feet diameter cooling water recirculation lines collapsed. No one was injured. However, the Bailly power plant was shut down for five months. Despite damage to the AFGD facility, and the congestion caused by having a major recovery effort on-site, construction of the AFGD facility was completed two weeks ahead of the original schedule. Start-up occurred on 2 June 1992, and commercial operations commenced on 15 June 1992.

The demonstration period will continue for three years, through 14 June 1995. During this period, six one-month demonstration tests will be performed, to assess scrubber operations with a variety of coals. All coals will be bituminous coals, with sulfur content ranging from 2.0% to 4.5%. The demonstration test scheduled is presented in Table 5.

Note that the first of these demonstration tests (Test No. 3), using the normal coal for the Bailly Station (3.0% to 3.5% sulfur), was successfully completed in September 1992. The second demonstration test (Test No. 4) using 3.5% - 4% Sulfur coal was completed in June 1993. Figures 2, 3, and 4 show the SO2 removal performance during this test at various Boiler Loads.

Summary

As of this report, the facility is exceeding all contractual requirements. The AFGD facility is removing in excess of 95% of the SO2 from Units #7 and #8, has a 99.9% availability rate, and is producing a wallboard-grade gypsum that is 98% pure.

Table 1. Wallboard-Grade Gypsum Specifications for Pure Air Scrubber at Bailly Station.

	Expected	<u>Achieved</u>
Gypsum Purity (wt. % dry)		
CaSO4 - 2 H2O	>93.0%	97.83%
CaSO3 - 1/2 H2O	<2.0%	0.082%
SiO2	<2.5%	0.37%
Fe2O3	<3.5%	0.21%
R2O3 (R= metal other than Fe)		0.29%
Chlorides	<120 ppm	23 ppm
Total Water Soluble Salts	<600 ppm	41 ppm
Free H2O (wt. %)	<10%	6.89%
Mean Particle Size (microns)	>20	43

Table 2. Water Requirements for Pure Air Scrubber at Bailly Station.

	Expected	Achieved
Supply Water Flow	<3,000 gpm	1,209 gpm
Process Wastewater Flow	<2,500 gpm	2,025 gpm
Wastewater pH	6.0 to 9.0	8.0 to 9.0
Wastewater Total	<30 ppm	<13 ppm
Suspended Solids Wastewater Dissolved Solids		
Chlorides (Cl)	<30,000 ppm	5,825 ppm
Sulfates (SO4 ⁺²)	<2,500 ppm	2,025 ppm
Fluorides (F)	<1,100 ppm	22 ppm
Calcium (Ca ⁺²)	<12,000 ppm	2,008 ppm
Magnesium (Mg ⁺²)	<6,000 ppm	1,483 ppm
Total	<100,000 ppm	16,025 ppm

Table 3. Operations Summary for Pure Air Scrubber at Bailly Station

	Expected	Achieved
SO2 Emissions	90% removal or 1.2 lb/MMBtu, whichever is less stringent	Averaged 95% (during DOE test up to 98+%, or 0.382 lb/MMBtu)
Power Consumption		
24-hour average	<8,650 kW	5,962 kW
instantaneous	<9,650 kW	6,128 kW
Facility Pressure Drop		
24-hour average	<13.5 IWC	6.66 IWC
instantaneous	<14.5 IWC	7.55 IWC
Particulate Emissions	no net increase	0.04 inlet
(g/SCFD)		0.0071 outlet
Facility Availability-Hrs.		99.97%
- MW	95%	99.96%
SO2 Removed (Tons)		79,248
Limestone Received	****	122,383
Gypsum Shipped (Wet)		216,344
Gypsum Moisture	<10%	6.63
Gypsum Chloride	<120 ppm	24
Gypsum Purity	93%	96.68
Average Water Consumption (GPM)	3,000	1,211
Average Waste Water Flow (GPM)	275	87

Table 4. AFGD Project Cost Summary

-	Budget	Actual/Estimate
Phase I (Design)	\$ 16,251,000	\$ 20,876,000
Phase II (Construction	\$ 93,142,000	\$ 85,654,000
Phase III (Operations	\$ <u>41.104.000</u>	\$ <u>43.067.000</u>
Sub-Total	\$150,497,000	\$149,597,000
PowerChipTM Gypsum	\$ <u>_1.210.898</u>	\$ <u>1.210.898</u>
Total	\$151,707,898	\$150,807,898

Table 5. AFGD Demonstration Test Schedule

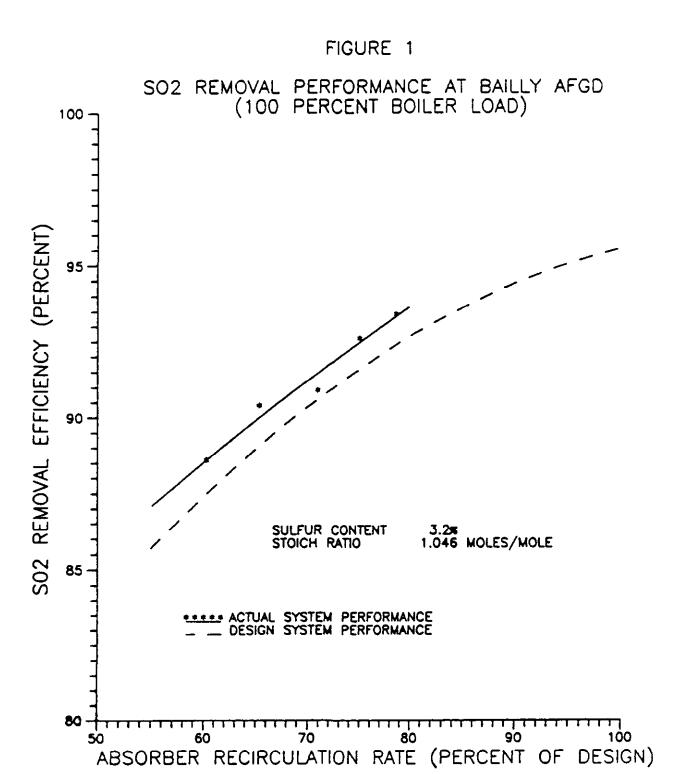
Test No.	Coal Sulfur	Schedule
1	2.0% to 2.5%	Fall 1994
2	2.5% to 3.0%	Fall 1993
3	3.0% to 3.5%	Fall 1992 (complete)
4	3.5% to 4.0%	Spring 1993 (complete)
5	4.0% to 4.5%	Spring 1994
6	Optimal Conditions	Spring 1995

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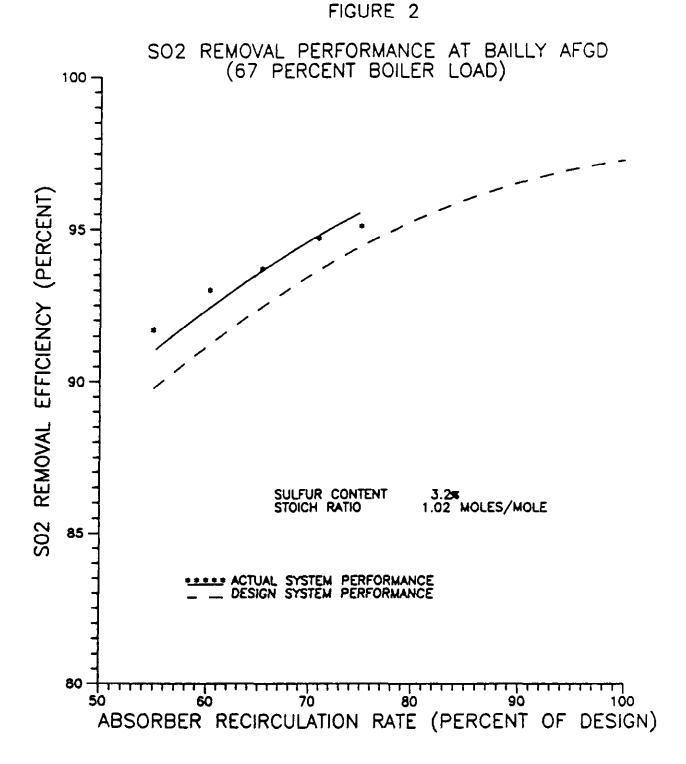
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PURE AIR'S ADVANCED FLUE GAS DESULFURIZATION CLEAN COAL TECHNOLOGY DEMONSTRATION PROJECT



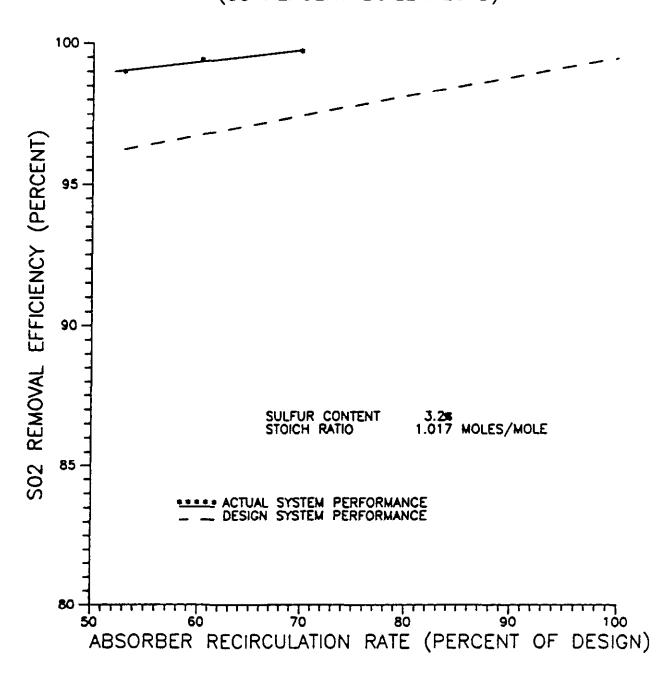
PURE AIR'S ADVANCED FLUE GAS DESULFURIZATION CLEAN COAL TECHNOLOGY DEMONSTRATION PROJECT



PURE AIR'S ADVANCED FLUE GAS DESULFURIZATION CLEAN COAL TECHNOLOGY DEMONSTRATION PROJECT

FIGURE 3

SO2 REMOVAL PERFORMANCE AT BAILLY AFGD
(33 PERCENT BOILER LOAD)



THE CLEAN COAL TECHNOLOGY PROGRAM 10 MWe DEMONSTRATION OF GAS SUSPENSION ABSORPTION FOR FLUE GAS DESULFURIZATION

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SECOND ANNUAL
CLEAN COAL TECHNOLOGY CONFERENCE
September 7-9, 1993
Atlanta, Georgia,

ABSTRACT

This paper presents a description and the initial test results of the Gas Suspension Absorption technology demonstration in the Clean Coal Technology project entitled "10 MW Demonstration of Gas Suspension Absorption." AirPol is currently performing this demonstration with the cooperation of the Tennessee Valley Authority under a Cooperative Agreement with the United States Department of Energy. This low-cost retrofit project seeks to demonstrate the Gas Suspension Absorption system, which is expected to remove more than 90% of the sulfur dioxide from coal-fired flue gas, while achieving a high utilization of reagent lime.

INTRODUCTION

AirPol, with the assistance of the Tennessee Valley Authority (TVA), is demonstrating the Gas Suspension Absorption (GSA) technology in the Clean Coal Technology project entitled "10 MW Demonstration of Gas Suspension Absorption." AirPol is performing this demonstration under a Cooperative Agreement awarded by the United States (U.S.) Department of Energy (DOE) in October 1990. This project was selected in Round III of the Clean Coal Technology Program.

This project is the first North American demonstration of the GSA system for flue gas desulfurization (FGD) for a coal-fired utility boiler. This low-cost retrofit project seeks to demonstrate the GSA system, which is expected to remove more than 90% of the sulfur dioxide (SO₂) from the flue gas, while achieving a high utilization of reagent lime. TVA has provided its National Center for Emissions Research (NCER) as the host site and is providing operation, maintenance, and technical support during the operations and testing phase of this project. The NCER is located at the TVA's Shawnee Fossil Plant near Paducah, Kentucky.

The experience gained by AirPol in designing, fabricating, and constructing the GSA equipment through the execution of this project will be used for future commercialization of the GSA technology. The results of the operation and testing phase will be used to further improve the GSA system design and operation.

The specific technical objectives of the GSA demonstration project are the following:

- Effectively demonstrate SO₂ removal in excess of 90% using high-sulfur U.S. coal.
- Optimize design and operating parameters to increase the SO₂ removal efficiency and the lime utilization.
- Compare the SO₂ removal efficiency of the GSA technology with existing spray dryer/electrostatic precipitator (SD/ESP) technology.

DOE has issued an amendment to the Cooperative Agreement to include the additional scope of work for air toxics testing and also the operation and testing of a 1 MWe pulse jet baghouse (PJBH) pilot plant in cooperation with TVA and the Electric Power Research Institute (EPRI). The two-fold purpose of this additional work is the following:

- Determine the air toxics removal performance of the GSA technology.
- Compare the SO₂, particulate, and air toxics removal performance between GSA/ESP and GSA/PJBH systems.

The PJBH can treat flue gas removed either upstream or downstream of the ESP. The testing of the PJBH will be conducted for both configurations.

The total budget for the project with the added scope of work is \$7,720,000; however, the project cost is currently under the budget. The favorable variance has resulted mainly from actual material and construction costs being much lower than the original estimate. The performance period of the project, including the air toxics measurements, PJBH testing, and report preparation is from November 1990 to February 1994.

AirPol began the design work on this project in November 1990, shortly after award of the Cooperative Agreement by DOE in October 1990. At the outset of the project, site access at the NCER was delayed by TVA to allow the completion of another project. That caused a one-year delay in this Clean Coal Technology project. The design phase of the GSA project was completed in December 1991. The fabrication and construction of the GSA unit was completed ahead of schedule in early September 1992. The planned one-year operation and testing of the demonstration unit began in late October 1992.

HISTORY OF THE GSA TECHNOLOGY

The GSA process is a novel concept for FGD that was developed by AirPol's parent company, F.L. Smidth miljo a/s in Copenhagen, Denmark. The process was initially developed as a

cyclone preheater system for cement kiln raw meal (limestone and clay). This innovative system provided both capital and energy savings by reducing the required length of the rotary kiln and lowering fuel consumption. The GSA system also showed superior heat and mass transfer characteristics and was subsequently used for the calcination of limestone, alumina, and dolomite. The GSA system for FGD applications was developed later by injecting lime slurry and the recycled solids into the bottom of the reactor to function as an acid gas absorber.

In 1985, a GSA pilot plant was built in Denmark to establish design parameters for SO₂ and hydrogen chloride (HCl) absorption for waste incineration applications. The first commercial GSA unit was installed at the KARA Waste-to-Energy Plant at Roskilde, Denmark, in 1988. Currently, there are ten GSA installations in Europe, and all are municipal solid waste incinerator applications.

With the increased emphasis on SO₂ emissions reduction by electric utility and industrial plants as required by the Clean Air Act Amendments of 1990, there is a need for a simple and economic FGD process, such as GSA, by the small to mid-size plants where a wet FGD system may not be feasible. The GSA FGD process, with commercial and technical advantages expected to be confirmed in this demonstration project, will be a viable alternative to meet the needs of the U.S. utility industry and the industrial boilers.

GSA FGD PROCESS DESCRIPTION

The GSA FGD system, as shown in the Figure 1 Process Flow Diagram, includes:

- A circulating fluidized bed reactor.
- A separating cyclone incorporating a system for recycling the separated material to the reactor.
- A lime slurry preparation system which proportions the slurry to the reactor via a dual-fluid nozzle.
- A dust collector which removes flyash and reaction products from the flue gas stream.

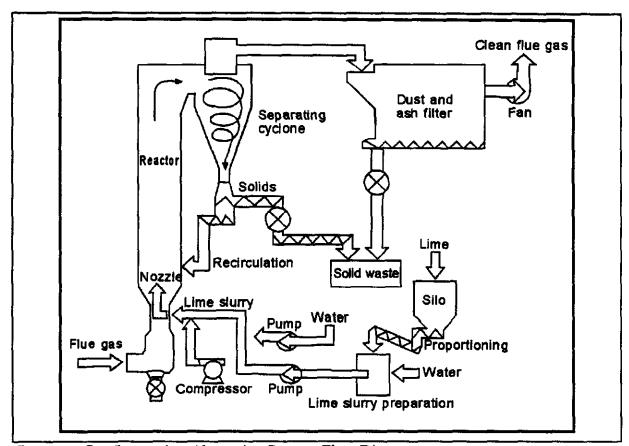


Figure 1. Gas Suspension Absorption Process Flow Diagram

The flue gas from the boiler air preheater is fed into the bottom of the circulating fluidized bed reactor where it is mixed with the suspended solids that have been wetted by the fresh lime slurry. The suspended solids consist of reaction products, residual lime, and flyash. During the drying process in the reactor, the moisture in the fresh lime slurry, which coats the outer surface of the suspended solids, evaporates. Simultaneously, the lime particles in the slurry undergo a chemical reaction with the acid components of the flue gas, SO₂ and HCl, capturing and neutralizing them.

The partially cleaned flue gas flows from the top of the reactor to the separating cyclone and then to an ESP (or a fabric filter), which removes the dust and ash particles. The flue gas, which has now been cleaned, is then released into the atmosphere through the stack.

The cyclone separates most of the solids from the flue gas stream. Approximately 95% to 99% of these collected solids are fed back to the reactor via a screw conveyor, while the remaining solids leave the system as a byproduct material. Some of these solids recirculated to the reactor are still reactive. This means that the recirculated lime is still available to react and neutralize the acid components in the flue gas.

The pebble lime is slaked in a conventional, off-the-shelf system. The resulting fresh slaked lime slurry is pumped to an interim storage tank and then to the dual-fluid nozzle. The slurry is diluted with trim water prior to being injected into the reactor.

Automatic Process Adjustment

An effective monitoring and control system automatically ensures that the required level of SO₂ removal is attained while keeping lime consumption to a minimum. This GSA control system, which is shown in Figure 2, incorporates three separate control loops:

- 1. Based on the flue gas flow rate entering the GSA system, the first loop continuously controls the flow rate of the recycled solids back to the reactor. The large surface area for reaction provided by these fluidized solids and the even distribution of the lime slurry in the reactor, provides for the efficient mixing of the lime with the flue gas. At the same time, the large volume of dry material prevents the slurry from adhering to the sides of the reactor.
- 2. The second control loop ensures that the flue gas is sufficiently cooled to optimize the absorption and reaction of the acid gases. This control of flue gas temperature is achieved by the injection of additional water along with the lime slurry. The amount of water added into the system is governed by the temperature of the flue gas exiting the reactor. This temperature is normally set a few degrees above flue gas saturation temperature to insure that the reactor solids will be dry so as to reduce any risk of acid condensation.

3. The third control loop determines the lime slurry addition rate. This is accomplished by continuously monitoring the SO₂ content in the outlet flue gas and comparing it with the required emission level. This control loop enables direct proportioning of lime slurry feed according to the monitored results and maintains a low level of lime consumption.

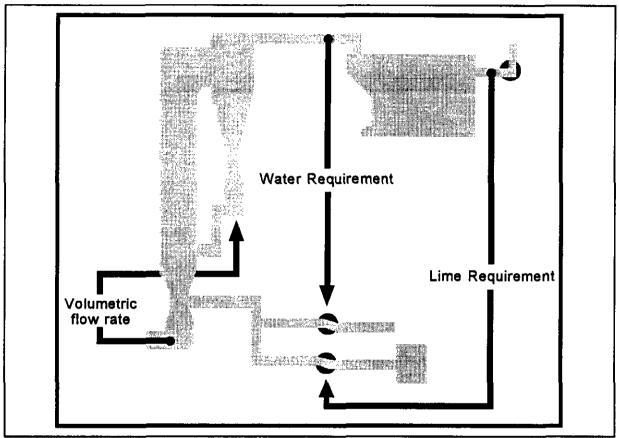


Figure 2. Gas Suspension Absorption Control System

COMPARISON OF GSA PROCESS WITH COMPETING TECHNOLOGY

Simplicity is the key feature of the GSA system. The advantages of the GSA system over competing technologies are summarized as follows:

Slurry Atomization

The major difference between the GSA and the competing technologies lies in how the reagent is introduced and used for SO₂ absorption. A conventional semi-dry scrubber:

- Requires a costly and sensitive high-speed rotary atomizer or a high-pressure atomizing nozzle for fine atomization,
- Absorbs SO₂ in an "umbrella" of finely atomized slurry with a droplet size of about 50 microns,
- May require multiple nozzle heads or rotary atomizer to ensure fine atomization and full coverage of the reactor cross section, and
- Uses recycle material in the feed slurry necessitating expensive abrasion-resistant materials in the atomizer(s).

The GSA process, on the other hand:

- Uses a low-pressure, dual-fluid nozzle,
- Absorbs SO₂ on the wetted surface of suspended solids with superior mass and heat transfer characteristics.
- Uses only one spray nozzle for the purpose of introducing slurry and water to the reactor, and
- Uses dry injection of recycle material directly into the reactor, thereby avoiding
 erosion problems in the nozzle or technical limitation on the amount of solids that
 can be recycled.

Simple and Direct Method of Lime/Solid Recirculation

The recirculation of used lime is the trend for semi-dry scrubbing systems. The recirculation of solids in the GSA system is accomplished using a feeder box under the cyclone, which introduces

the material directly into the reactor. The recirculation feature commonly used in most other semi-dry processes has an elaborate ash handling system to convey and store the ash. The method of introducing the recirculated material is usually by mixing it with the fresh lime slurry. The presence of ash in the lime slurry may cause sediment problem in the slurry lines and excessive nozzle wear.

High Acid Gas Absorption

The GSA reactor is capable of supporting an extremely high concentration of solids (recirculated material) inside the reactor, which acts like a fluidized bed. This concentration will normally be as high as 200-800 grains/scf. These suspended solids provide a large surface area for contact between the lime slurry (on the surface of the solids) and the acidic components in the flue gas. This high contact area allows the GSA process to achieve levels of performance that are closer to that of a wet scrubber, rather than a dry scrubber. Since drying of the solids is also greatly enhanced by the characteristic large surface area of the fluidized bed, the temperature inside the reactor can be reduced below that of the typical semi-dry scrubber. This lower operating temperature facilitates the acid gas removal in the GSA system and helps it achieve SO₂ removal levels which are comparable to a wet scrubber.

Low Lime Consumption / Minimum Waste Byproduct Residue

The design of the GSA reactor allows for more efficient utilization of the lime slurry because of the high internal recirculation rate and precise process control. The higher lime utilization (up to 80%) lowers the lime consumption, thereby minimizing one of the major operating costs. In addition, the lower lime consumption reduces the amount of byproduct generated by the system.

Low Maintenance Operation

Unlike the typical semi-dry scrubbers, the GSA system has no moving parts inside the reactor, thus ensuring relatively continuous, maintenance-free operation. The orifice diameter of the GSA

injection nozzle is much larger than that used in a conventional semi-dry process, and there is little chance for it to plug. Nozzle wear is also minimized. Should the need for replacing the nozzle arise, it can be replaced in a few minutes. The cyclone also has no moving parts. Both the reactor and the cyclone are fabricated from unlined carbon steel.

The GSA process also has few pieces of equipment. Most of the equipment is in the lime slurry preparation area, which typically is an off-the-shelf item, and the technology is well known.

No Internal Buildup

By virtue of the fluidized bed inside the reactor, the inside surface of the reactor is continuously "brushed" by the suspended solids and is kept free of any buildup. Internal wall buildup can be a problem with the conventional semi-dry scrubber. There is also no wet/dry interface on any part of the equipment and this avoids any serious corrosion problem.

Modest Space Requirements

Due to the high concentration of suspended solids in the reactor, more than adequate reaction occurs in a relatively short period of time. A high flue gas velocity of 20 to 22 feet per second as compared to 4 to 6 feet per second for a semi-dry scrubber and the shorter residence time of 2.5 seconds as compared to 10 to 12 seconds for a semi-dry scrubber, allow for a smaller diameter reactor which leads to a considerable reduction in space requirements.

Short Construction Period

The compact design of the GSA unit requires less manpower and time to be erected as compared to the typical semi-dry scrubbers. Despite the relatively complicated tie-ins and extremely constrained work space, the retrofit GSA demonstration unit at the TVA's NCER was erected in three and a half months.

Heavy Metals Removal

Recent test results from waste incineration plants in Denmark indicate that the GSA process is not only effective in removing acidic components from the flue gas but is also capable of removing heavy metals, such as mercury, cadmium, and lead. This heavy metal removal capability of the GSA process at the NCER will be evaluated later in the test program.

PROJECT STATUS AND KEY MILESTONES

The project schedule and tasks involved in the design, construction, and operation and testing phases are as follows:

Phase I - En	gineering and Design	Start - End
1.1	Project and Contract Management	11/01/90-12/31/91
1.2	Process Design	11/01/90-12/31/91
1.3	Environmental Analysis	11/01/90-12/31/91
1.4	Engineering Design	11/01/90-12/31/91
Phase II - Pr	ocurement and Construction	
2.1	Project and Contract Management	01/01/92-09/30/92
2.2	Procurement and Furnish Material	01/01/92-04/30/92
2.3	Construction and Commissioning	05/01/92-09/30/92
Phase III - C	Operating and Testing	
3.1	Project Management	10/01/92-02/28/94
3.2	Start-up and Training	10/01/92-10/14/92
3.3	Testing and Reporting	10/15/92-02/28/94

The progress of this project has been on or ahead of schedule. The parametric optimizing tests are scheduled for completion in August 1993. Following the air toxics testing, which is scheduled to be conducted in September, there will be a one month around-the-clock demonstration run.

TEST PLAN

A test plan was prepared to depict in detail the procedures, locations, and analytical methods to be used in the tests. The following objectives are expected to be achieved by testing the GSA system:

- Optimization of the operating variables.
- Determination of stoichiometric ratios for various SO₂ removal efficiencies.
- Evaluation of erosion and corrosion at various locations in the system.
- Demonstration of 90% or greater SO₂ removal efficiency when the boiler is fired with high-sulfur coal.
- Determination of the air toxics removal performance.
- Evaluation of the PJBH performance in conjunction with the GSA process.

Optimization Tests

The optimization of the SO₂ removal efficiency in the GSA system will be accomplished through the completion of a statistically-designed factorial test plan. For each test series, the GSA system is set to operate at a certain combination of operating parameters. The results of these test series are analyzed statistically to determine the impact of the operating parameters, thus arriving at the optimum operating point for the GSA process at the various operating conditions expected in future applications. Operating parameters that may be varied in different test series for process optimization purposes are the following:

- Inlet flue gas flow rate
- Inlet SO₂ concentration (dependent on availability of different coal)
- Inlet flue gas temperature
- Inlet dust loading
- Solids recirculation rate
- Stoichiometric ratio
- Approach-to-saturation temperature
- Coal chloride level

Preliminary tests were conducted prior to the factorial testing to determine the ranges that these parameters could be varied. The results from the preliminary testing were used as the basis for the design of the factorial test program.

Data Collection

The following data will be sampled and recorded during the tests by either the computerized data sampling and recording system (via field mounted instruments) or manual field determinations:

- Inlet flue gas flow into the system
- SO₂ and HCl loading at the system inlet, SO₂ loading at the ESP inlet and outlet
- · Flue gas temperature at the system inlet, the reactor outlet, and the ESP outlet
- Particulate loading at the ESP inlet and outlet
- Fresh lime slurry flow rate and composition (for lime stoichiometry calculation)
- Water flow rate
- Wet-bulb temperature at the reactor inlet (for approach-to-saturation temperature calculation)
- Coal analysis (proximate and ultimate)
- Lime analysis
- Byproduct rate and composition
- Water analysis
- Power consumption

Preliminary Testing

Immediately after the dedication of the AirPol GSA demonstration plant in late October 1992, a series of preliminary tests was begun. The purpose of these tests was to investigate the operating limits of the GSA system as installed at the NCER. The results from several of the preliminary tests completed at the NCER in November and December were very interesting. During one of these tests, the approach-to-saturation temperature in the reactor was gradually decreased and the overall system (reactor/cyclone and ESP) SO₂ removal efficiency was monitored over this four-day test. The overall system SO₂ removal efficiency increased from about 65% to more than 99% at the closest approach-to-saturation temperature (5°F). The other conditions, which remained constant, were 320°F inlet flue gas temperature, 1.40 moles

Ca(OH)₂/mole inlet SO₂ for the lime stoichiometry, and essentially no chloride in the system. The SO₂ removal results from this test are shown in Figure 3.

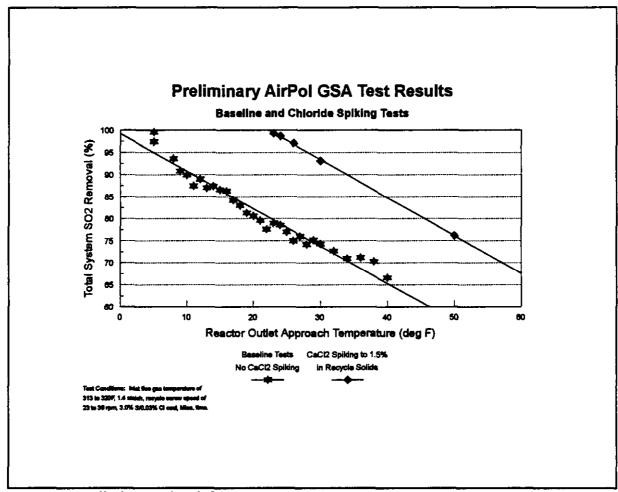


Figure 3. Preliminary AirPol GSA Test Results

The data from this test show that the SO₂ removal efficiency increased dramatically as the flue gas temperature in the reactor more closely approached the saturation temperature of the flue gas, with the incremental increases in the SO₂ removal becoming more and more significant as the approach-to-saturation temperature declined. The ability of the GSA system to operate at this close approach-to-saturation temperature without any indication of plugging problems was surprising. Later analysis showed that the moisture level in the solids remained below 1%.

A second extended test was run during December. This test was run at the same conditions as the previous test, except that in this test, calcium chloride was added to the system to simulate the combustion of a high-chloride (about 0.3%) coal. Previous work by TVA at the NCER had demonstrated that spiking these semi-dry, lime-based FGD processes with a calcium chloride solution adequately simulated a high chloride coal application. Again, the approach-to-saturation temperature was gradually decreased over a four-day period with all other conditions held constant and the overall system SO₂ removal efficiency was monitored. The preliminary results from this second test are also shown in Figure 3 above.

The overall system SO₂ removal efficiency increased from about 70% at the high approach-to-saturation condition to essentially 100% at the closer approach-to-saturation temperature (23°F). No attempt was made to operate the system at the close approach-to-saturation temperatures used in the first test because the SO₂ removal efficiency was approaching 100%. In addition, there were initially some concerns about the secondary effect of calcium chloride addition. Calcium chloride is an ionic salt that tends to depress the vapor pressure of water in the system and thus, slows the evaporation of water from the slurry. Calcium chloride is also a hygroscopic material, which means it has the ability to absorb moisture from the humid flue gas. The increased moisture in the "dry" solids allows more reaction with SO₂, but also increases the potential for plugging in the system. The easiest method for mitigating this potential for plugging is to increase the approach-to-saturation temperature in the reactor. However, the moisture levels in the solids during this test remained below 1%, even at the closest approach-to-saturation temperature.

Another interesting finding from the preliminary testing is that the GSA process is capable of supporting a very high level of recirculation material in the reactor. This high solid concentration inside the reactor is the reason for the superior drying characteristics of the GSA system. Based on the results from these initial tests, the recycle rate back to the reactor was doubled prior to starting the factorial testing.

Factorial Testing

The primary focus of the recent GSA testing was the completion of the statistically-designed factorial test program. The purpose of this factorial testing was to determine the effect of the process variables on the SO₂ removal efficiency in the reactor/cyclone and the ESP.

Based on the successful preliminary testing, the major process design variables were determined, levels for each of these variables were defined, and an overall test plan was prepared. The major variables were approach-to-saturation temperature, lime stoichiometry, flyash loading, coal chloride level, flue gas flow rate, and recycle screw speed. Two levels were determined for nearly all of the variables and these variables and levels are shown in the table below. The one exception was the approach-to-saturation temperature where three levels were defined, but the third level was only run for those tests at the lower coal chloride level.

Major Variables and Levels for Factorial Testing Table		
Variable		Level
Approach-to-saturation temperature	°F	8*, 18, and 28
Ca/S	moles Ca(OH) ₂ /mole inlet SO ₂	1.00 and 1.30
Flyash loading	gr/acf	0.5 and 2.0
Coal chloride level	%	0.02 and 0.12
Flue gas flow rate	kscfm	14 and 20
Recycle screw speed	rpm	30 and 45
*8°F level run only at the low-chloride level		

Table 1. Major Variables and Levels for Factorial Testing

Although the preliminary chloride spiking tests had not been run at an approach-to-saturation temperature below 23°F, the decision was made to complete these chloride-spiking factorial tests at an 18°F approach-to-saturation temperature. There was some risk in this decision because the water evaporation rate is decreased at the higher chloride levels. However, based on previous test work at the NCER, the expectation was that at the lower chloride levels in this test plan, equivalent to a coal chloride level at 0.12%, the GSA system could operate at the 18°F approach-to-saturation temperature condition.

RESULTS OF FACTORIAL TESTING

SO, Removal Efficiency

The overall system SO₂ removal efficiency results from these factorial tests are currently being analyzed, but several general relationships have become apparent. First, as was expected based on the previous testing at the NCER, significant positive effects on the SO₂ removal efficiency in the system came from increasing the lime stoichiometry and other factors such as increasing the coal chloride level or decreasing the approach-to-saturation temperature. Increasing the recycle rate resulted in higher SO₂ removal, but the benefit appeared to reach an optimum level, above which further increases in the recycle rate did not seem to have a significant effect on SO₂ removal. Increasing the flue gas flow rate had a negative effect on the SO₂ removal in the system.

The overall system SO₂ removal efficiency during these tests ranged from slightly more than 60% to nearly 95%, depending on the specific test conditions. The higher SO₂ removal efficiency levels were achieved at the closer approach-to-saturation temperatures (8 and 18°F), higher lime stoichiometry level (1.30 moles Ca(OH)₂/mole inlet SO₂), and the higher coal chloride level (0.12%). The lower SO₂ removal efficiency levels were achieved at the higher approach-to-saturation temperature (28°F), the lower lime stoichiometry level (1.00 mole Ca(OH₂/mole inlet SO₂), and the lower coal chloride level (0.02-0.04%). Most of the SO₂ removal in the GSA system occurs in the reactor/cyclone, with only about 2 to 5 percentage points of the overall

system removal occurring in the ESP. There is substantially less SO₂ removal in the ESP than in the previous testing at the NCER, but the overall system SO₂ removal efficiencies appear to be higher with the GSA process for most test conditions.

As one would expect, the lime stoichiometry level, which was tested at 1.00 and 1.30 moles Ca(OH)₂/mole inlet SO₂, seems to have the most significant effect on the SO₂ removal efficiency in the GSA system.

The approach-to-saturation temperature, which was evaluated at three levels of 8, 18, and 28°F for the low coal chloride conditions and the two levels of 18 and 28°F for the higher coal chloride condition, appears to be the second most important variable in the GSA system in terms of the overall system SO₂ removal efficiency.

The third most important variable seems to be the chloride level in the system. Two coal chloride levels were tested, the baseline coal chloride level of 0.02 to 0.04% and the equivalent of a 0.12% coal chloride level. The higher chloride level was achieved by spiking the feed slurry with a calcium chloride solution.

One of the most surprising results of this factorial testing was the ability of the GSA system to operate at an 8°F approach-to saturation temperature at the low-chloride condition without any indication of plugging. This is even more impressive given the very low flue gas residence time in the reactor/cyclone. The second interesting result of this testing was the ability of the GSA system to operate at the 18°F approach-to-saturation temperature at the higher chloride level. In the preliminary testing at a much higher coal chloride level (0.3%), the lowest approach-to-saturation temperature tested was 23°F. No operating problems were encountered in the tests completed at the 0.12% coal chloride level and 18°F approach-to-saturation temperature conditions. In fact, the average moisture level in the solids remained below 1.0% in all of these factorial tests, even at the higher coal chloride level.

ESP Performance

The ESP installed at the NCER is a relatively modern, 4-field unit with 10 in. plate spacing, similar in design to several full-scale ESPs installed on the TVA Power System. This unit has 23-ft.-high plates with 8 parallel gas passages. The specific collection area (SCA) of this unit is about 440 ft²/kacfm under the cooled, humidified flue gas conditions downstream of the reactor/cyclone. (For the untreated flue gas at 300°F, i.e., in a fly-ash-only application, the SCA of this ESP is about 360 ft²/kacfm.)

The particulate removal performance of this ESP was determined for each of the factorial tests, even though this was not the primary focus of the testing. The most important result of this particulate testing was that the emission rate from the ESP was substantially below the New Source Performance Standards (NSPS) for particulate (0.03 lb/MBtu) at all of the test conditions evaluated. The typical emission rate was 0.010 lb/MBtu. The particulate removal efficiency in the ESP for nearly all of the tests was above 99.9% and the outlet grain loadings were below 0.005 gr/acf.

However, during the testing there were disturbing indications of low power levels in the first field of the ESP, particularly in those tests involving chloride spiking. In some of these chloride-spiking tests completed at the high flue gas flow rate (20,000 scfm), the power level in the first field was only about 5% of the normal level, effectively meaning that the first field had "collapsed." Even with these low power levels in the first field of the ESP, the particulate removal efficiencies were still 99.9+% and the emission rate was in the range of 0.010 lb/MBtu. The cause of these low power levels in the first field of the ESP is being investigated. These low power levels could be the result of a number of factors, including plate-wire alignment problems as observed in a recent internal inspection.

One surprising result of this ESP testing was that there was no significant improvement in the ESP performance with increasing SCA. For some of these tests, the SCA in the ESP approached

800 ft²/kacfm and the flue gas velocity in the ESP dropped below 2.0 ft/sec and yet the emission rate remained in the same range as in the other tests, i.e., 0.010 lb/MBtu.

Pulse Jet Baghouse Performance

Although not part of the original GSA project, TVA and EPRI had cofunded the installation of a 1-MWe PJBH pilot plant at the NCER to be operated in conjunction with the existing GSA demonstration. Later, AirPol and DOE joined in the operation and testing of this PJBH pilot plant program. The PJBH pilot plant, which was started up in late January, can pull a slipstream of flue gas from either the ESP inlet or outlet, as shown in Figure 1. In the first series of factorial tests, the PJBH pilot plant pulled flue gas from the ESP inlet and thus, treated flue gas with the full particulate loading (3 to 5 gr/acf) from the GSA reactor/cyclone. The inlet flue gas flow rate was about 5,000 acfm, which corresponds to an air-to-cloth ratio (A/C) of 4.0 acfm/ft² in the PJBH. During the second series of factorial tests, the PJBH pilot plant pulled flue gas from the ESP outlet. The same inlet flue gas flow rate was treated (5,000 acfm), but two-thirds of the bags were removed prior to this testing and thus, the A/C for these tests was 12 acfm/ft².

The cleaning of the bags in the PJBH was pressure-drop-initiated during this testing with the cleaning cycle beginning whenever the tubesheet pressure drop reached 6 in. of water. The cleaning continued until the tubesheet pressure drop had declined to about 4-1/2 inches of water. The bags were cleaned by a low-pressure, high-volume, ambient air stream delivered by a rotating manifold.

SO, Removal Efficiency

The SO₂ removal efficiency in the reactor/cyclone/PJBH system was typically about 3-5 percentage points higher than that achieved in the reactor/cyclone/ESP system at the same test conditions. This higher SO₂ removal efficiency in the PJBH system was not unexpected given the intimate contact between the SO₂-laden flue gas and the solids collected on the outside of the bags as the flue gas passed through the filter cake and the bags before being discharged to the

stack. However, it should be noted that most of the SO₂ removal occurred in the reactor/cyclone and the PJBH SO₂ removal efficiency, based on the inlet SO₂ to the reactor, contributed less than 8 percentage points to the overall system SO₂ removal efficiency during this testing.

Particulate Removal

The particulate removal efficiency in the PJBH was 99.9+% for all of the tests completed with the full dust loading from the GSA reactor/cyclone. The emission rate for all of these tests was well below the New Source Performance Standards for particulates and was typically in the range of 0.010 lb/MBtu.

Demonstration Run

Based on the findings during the factorial testing, the GSA system will be operated at optimum settings for a four-week consecutive period of around-the-clock operation to demonstrate the reliability of the system operation as well as its SO₂ removal capability. During the demonstration run, all controls will be switched to automatic mode with set points determined from the optimizing tests.

COMMERCIALIZATION

One of the objectives of this demonstration project is for AirPol to establish its capability in designing, fabricating, and constructing the GSA system so that the demonstrated technology can be effectively commercialized for the benefit of the U.S. electric utility and industrial markets. The progress of this demonstration project matches very well with the development of the utility FGD market. The GSA technology is now ready to be commercialized for the industry in order to meet the Phase II Clean Air Act Amendments compliance requirements.

During the course of designing the demonstration unit, an effort was made by AirPol to standardize the process design, equipment sizing, and detailed design so that the installation of

a commercial unit can be accomplished within a relatively short time frame. An effort was also made during the design phase to achieve simplicity in the equipment design, which later proved to contribute to reduced material and construction costs. Another major effort being made at AirPol now is to scale up the GSA design to accommodate a utility plant up to 200 MWe with a single GSA reactor. Having gained the confidence that the GSA system is capable of achieving the required levels of performance, the current effort being made at AirPol is to develop standard design of scale-up units. Meanwhile, field operating experience and findings continue to help perfect the design process.

DISCUSSION

As of September 1993, the design, fabrication, installation, and performance optimization of the GSA system for the Clean Coal Technology demonstration project will have been completed. AirPol will have successfully demonstrated the technical performance of the GSA FGD process.

It is expected that the results of the air toxics test will confirm that GSA is also capable of removing heavy metals such as mercury, cadmium, and lead. As this demonstration program is coming close to its completion, it can be concluded now that the GSA process is a viable solution to the SO₂ removal problem of coal-fired boiler plants, and AirPol is ready to offer the technology for commercial application.

DISCLAIMER

Reference in this report to any specific commercial product, process, or service is to facilitate understanding and does not necessarily imply its endorsement or favoring by either DOE or TVA.

LIFAC SORBENT INJECTION FOR

FLUE GAS DESULFURIZATION

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Abstract

This paper discusses the demonstration of LIFAC sorbent injection technology at Richmond Power and Light's (RP&L) Whitewater Valley Unit No. 2 under the auspices of the U.S. Department of Energy's (DOE) Clean Coal program. LIFAC is a sorbent injection technology capable of removing 75 to 85 percent of a powerplant's SO₂ emissions using limestone at calcium to sulfur molar ratios of between 2 and 2.5 to 1. The site of the demonstration is a coal-fired electric utility powerplant located in Richmond, Indiana, which is between Indianapolis, Indiana and Dayton, Ohio. The project is being conducted by LIFAC North America, a joint venture partnership of Tampella Power Corporation and ICF Kaiser Engineers, in cooperation with DOE, RP&L, and several other organizations including the Electric Power Research Institute (EPRI), the State of Indiana, and Black Beauty Coal Company.

Introduction

The Clean Coal Technology Program (CCT) has been recognized in the National Energy Strategy as a major initiative whereby coal will be able to reach its full potential as a source of energy for the nation and the international marketplace. Attainment of this goal depends upon the development of highly efficient, environmentally sound, competitive coal utilization technologies responsive to diverse energy markets and varied consumer needs. The CCT Program is an effort jointly funded by government and industry whereby the most promising of the advanced coal-based technologies are being moved into the marketplace through demonstration. The CCT Program is being implemented through a total of five competitive solicitations. This paper discusses the LIFAC sorbent injection technology which was selected in the third round of CCT solicitations.

LIFAC North America, a joint venture partnership of ICF Kaiser Engineers, Inc. and Tampella Power Corporation of Finland, will demonstrate the LIFAC flue gas desulfurization technology developed by Tampella. This technology provides sulfur dioxide emission control for powerplants, especially existing facilities with tight space limitations. Sulfur dioxide emissions are expected to be reduced by up to 85% by using limestone as a sorbent. The limestone is injected into the upper regions of a furnace, where calcining to lime and partial absorption of SO₂ occur. Subsequently, the combustion gas is passed through a unique piece of equipment known as the activation reactor. This is a vertical elongation of the ductwork between the air preheater and ESP where the combustion gas is humidified and SO₂ absorption is completed. The LIFAC technology will be demonstrated at Whitewater Valley Unit No. 2, a 60-MWe coal-fired powerplant owned and operated by Richmond Power and Light (RP&L) and located in Richmond, Indiana. The Whitewater plant consumes high-sulfur coals with sulfur contents ranging from 2.0 - 2.9 percent.

The project, co-funded by LIFAC North America and DOE, is being conducted with the participation of Richmond Power and Light, the State of Indiana, the Electric Power Research Institute, and the Black Beauty Coal Company. The project has a total cost of 21.4 million dollars and a duration of 48 months from the preliminary design phase through the testing program.

The sponsors of this project believe that LIFAC has the potential to be a new and important SO_2 control option for U.S. utilities subject to the Clean Air Act's acid rain regulations. To be considered as a commercially feasible option in this particular emissions control market, LIFAC must demonstrate a high SO_2 removal rate while remaining competitive with other options on a cost per ton of SO_2 removed basis. To this end, the sponsors of this project have designed the demonstration with the following goals in mind:

- Sustained High SO₂ Removal Rate Incorporated into the test plan are several periods of long term testing which are intended to demonstrate LIFAC's SO₂ removal and reliability characteristics under normal operating conditions.
- Cost LIFAC must compete with both low capital cost, low SO₂ removal rate options such as sorbent injection and high capital cost, high SO₂ removal rate options such as wet

scrubbing. This project will demonstrate LIFAC's competitiveness on a cost per ton of SO₂ removed basis with these currently available alternatives.

- Retrofit Adaptability The host site chosen required a retrofit with tight construction conditions that will prove LIFAC's ability to be installed where other technologies might not be possible. Construction was also to demonstrate LIFAC's ability to be built and brought on-line with zero plant down time other than scheduled outages.
- System Compatibility A major concern of utilities is the degree of compatibility of SO₂ removal systems with their existing operations. This demonstration will show LIFAC's minimal impact on the host site's boiler and associated subsystems.

LIFAC Process History and Description

In 1983, Finland enacted acid rain legislation which applied limits on SO₂ emissions sufficient to require that flue gas desulfurization systems have the capability to remove about 80 percent of the sulfur dioxide in the flue gas. This level could be met by conventional wet limestone scrubbers but not by then available sorbent injection technology. Tampella, therefore, began developing an alternative sorbent injection system which resulted in the LIFAC process.

Initially, development first involved laboratory and pilot plant tests, then full-scale tests of sorbent injection of limestone. Using high-ash, low-sulfur coal and a Ca/S molar ratio of three to one, Tampella was unable to achieve a 50 percent SO₂ removal rate at it's 160 megawatt Inkeroinen facility. Substitution of lime for limestone was rejected due to its high cost.

Subsequent research and development by Tampella led to the addition of a humidification section after the furnace which became known as the LIFAC process. The sorbent injection process was installed full scale on a 220 megawatt boiler located at Kristiinankaupunki, Finland and a side-stream representing 2.5 megawatts was used to test a small scale humidification reactor. SO₂ removal rates of up to 84 percent were achieved at this plant. Additional tests at the Neste Kulloo combustion laboratory were conducted at 8 megawatts and also achieved 84 percent removal rates.

In 1986, the first large full scale test was performed at Imatran Voima's Inkoo powerplant using a 70 megawatt side-stream from a 250 megawatt boiler. A 76 percent SO₂ removal rate with 1.5% sulfur coal was reached. A second LIFAC activation reactor was constructed to handle an additional 125 megawatt side-stream. This newer reactor is achieving removal rates of 75 to 80 percent while using Ca/S molar ratios of between 2 and 2.5 to 1. Also, in 1988 the first tests with high-sulfur U.S. coals were run at the Neste Kulloo Laboratory. A Pittsburgh No. 8 Seam coal containing 3 percent sulfur was tested and an SO₂ removal rate of 77 percent was achieved at a Ca/S molar ratio of 2 to 1.

LIFAC Process Description

The LIFAC system combines conventional limestone injection into the upper furnace region with a post-furnace humidification reactor located between the air preheater and the ESP. The process produces a dry, stable waste product that is removed from both the bottom of the humidification reactor and the ESP.

Finely pulverized limestone is pneumatically conveyed and injected into the upper region of the boiler where temperatures are approximately 1800 to 2200 degrees Fahrenheit. At these temperatures the limestone (CaCO₃) calcines to form lime (CaO) which readily reacts with the SO₂ to form calcium sulfate (CaSO₄). All of the sulfur trioxide (SO₃) reacts with the CaO to form CaSO₄.

Approximately 25 percent of the sulfur dioxide removal occurs in the boiler with the remaining 75 percent and the unreacted lime passing through the air preheater to the humidification reactor. There the flue gas is sprayed with water that allows the unreacted lime to hydrate to Ca(OH)₂ which more readily reacts with the sulfur dioxide and forms CaSO₃. A combination of the proper water droplet size and residence time allows for effective hydration of the lime and complete water evaporation to create a dry reactor bottom product.

After exiting the humidification reactor, the flue gas is reheated before entering the ESP. The humidification and lower gas temperature enhance the efficiency of the ESP. Seventy-five percent of the LIFAC-produced spent sorbent and fly ash is collected by the ESP with the other 25 percent collected by the humidification reactor. Both the reactor and ESP ash may be recycled to a point ahead of the reactor to improve sorbent utilization and to improve the SO₂ removal efficiency of the system to the range of 75 to 85 percent. A schematic of the LIFAC process is shown in Figure 1 along with the typical sampling locations used during the demonstration.

Process Advantages

LIFAC is similar to other current sorbent injection technologies but has unique advantages with its use of a patented vertical humidification reactor. And while LIFAC's sulfur dioxide removal efficiency is not as high as traditional wet flue gas desulfurization systems, its cost and simplicity of design, construction and operation offer other advantages over these alternative systems. In particular the advantages of the LIFAC system are:

- High SO₂ removal rates Currently available sorbent injection systems have been unable to sustain high SO₂ removal rates with any consistency. LIFAC has proven in the past and intends to demonstrate during this project the ability to achieve and sustain high SO₂ removal rates of 75 to 85 percent over long operating periods.
- By-products Wet lime and limestone scrubbing systems create a wet byproduct ash that must be further treated before disposal. LIFAC produces a dry solid waste ash

containing calcium sulfide, calcium sulfate and fly ash. This waste is easily disposed of under U.S. regulatory requirements, may be recycled to increase LIFAC's efficiency and may have commercial applications in the cement industry.

• Compatibility and Adaptability - LIFAC has minimal impact on the host's site and systems, primarily the boiler, ESP and ID fan. In addition, LIFAC requires little space and few utilities and therefore is easily installed even in small or cramped powerplant sites.

Construction and Systems Integration

Construction of the LIFAC system has occurred in two phases over a period of one and a half years. The first phase of construction was completed during a routine plant outage in March, 1991. The period was utilized to install tie-ins to the host site's existing systems.

Ductwork and three dampers were installed between the air preheater and ESP to allow flue gas flow to the LIFAC activation reactor. Tie-ins were also made to the powerplant's high-pressure steam, condensate and river-water supplies. The high-pressure steam is required to reheat the flue gas exiting the LIFAC reactor and the water is needed for flue gas humidification inside the reactor. Injection ports were also installed in the boiler walls about 10 feet above the nose elevation.

The second phase of construction began in the Fall of 1991 with the driving of reactor piling and the installation of underground conduit runs. Work continued through to the Summer of 1992 with no need for plant downtime other than normally scheduled outages. During this time the limestone storage area was completed and the injection system was installed on Unit #2. The activation reactor was constructed and then tested with both cold air during a scheduled Unit #2 outage and hot flue gas during a low electricity demand period. Other powerplant tie-ins such as the steam and condensate system were also tested during low demand periods in the evening or on weekends.

All of the construction work associated with the LIFAC system was performed in close proximity to the exterior of the powerplant or in cramped areas inside the plant. The ductwork tie-ins and new steelwork required inside the plant are located in small, difficult to access work areas. The reactor structure is approximately ten feet from the powerplant with the outside ductwork and piping crossing overtop of offices and the plant maintenance area. All of these new structures and equipment were constructed with no interference to daily plant operations.

Schedule

The current schedule for the LIFAC demonstration program extends over a four year period from the beginning of preliminary design in August 1990 through the testing program to be completed in early August 1994 (see Figure 2). The LIFAC system was originally scheduled

to come on-line in June of 1992 but due to delays in receiving construction permits and some minor startup problems, this date was moved to March 1993. Although testing is scheduled to continue through the Summer of 1994, preliminary test results are now available.

Currently the demonstration project is on track with the revised schedule shown in Figure 2. All construction work was completed at the beginning of August 1992. Equipment check-out was performed in July and August and the first limestone delivery was received in early September. Initial tests with limestone injection into the boiler along with post-furnace humidification were conducted in October to December 1992. Having overcome all the normal operational problems that accompany retrofit installations, the project team was prepared to conduct the test plan beginning in early 1993.

Test Plan

The test plan for the LIFAC demonstration is composed of five distinct phases, each with its own objective. The first of these phases, which has already been completed, consisted of the initial baseline testing portion of the project. Measurements were taken to characterize the operation of the host's boiler and associated subsystems prior to the use of the LIFAC system. The results will be used for comparison purposes with the LIFAC system in operation and with data collected at the end of the project to determine any changes in the host's systems.

The second, or parametric, phase of testing is currently underway and will be performed to determine the best combination of LIFAC process variables for SO₂ removal. The variables being studied include the limestone injection nozzles' angle and location, the Ca/S molar ratio, the need for supplemental injection air at the boiler, the water droplet size and injection nozzle arrangement in the reactor, the ash recycling ratio and the approach to saturation temperature of the flue gas exiting the activation reactor. The best combination of these variables will be chosen at the conclusion of this phase and used for the remainder of the test program.

Optimization tests will be conducted to examine the effects of different coal and limestone feeds on the SO₂ capture rate. Coals with sulfur contents as high as 3.3 percent will be tested to determine LIFAC's compatibility with high sulfur U.S. coals. Limestones with different compositions will also be tested to determine the LIFAC system's adaptability to local sorbent sources.

Long term testing will also be performed to demonstrate LIFAC's performance under commercial conditions. The LIFAC system will be in operation 24 hours per day for several weeks using the powerplant's baseline coal, high calcium limestone and the optimum combination of process variables. In addition to process performance measurements, during this phase the operation and maintenance requirements of the system will be examined. Long term (two to three weeks) tests will also be conducted with two other coals; one lower sulfur coal (1.5%) and one higher sulfur content coal (3.3%).

The final phase of testing is composed of the post-LIFAC tests. The baseline tests will be repeated to gather information on the condition of the boiler and its associated subsystems. Comparisons will be made to the original baseline data to identify any changes either caused by the LIFAC system or independent of its operation.

Preliminary Results

Once startup and checkout were complete and the operational problems overcome, the project team initiated parametric testing. During the early tests with limestone addition and humidification, increases in opacity levels prevented the system from being operated as intended. Test work conducted by EPRI and Southern Research Institute identified the cause as lower ash resistivity resulting from reduced operating temperatures in the ESP. The activation reactor is designed to humidify the flue gas and drop the gas temperature to slightly above saturation temperature in order to maximize SO₂ capture. The flue gas is then reheated to above 175° Fahrenheit as it exits the activation reactor. Due to the relatively small size of the ESP (only 200 SCA) and because of lower ash resistivities, it was determined the ESP needed to be operated at about 200° Fahrenheit to avoid any problems with increased opacity. Having determined this, the operating procedures were revised to insure an ESP operating temperature above 200° Fahrenheit.

Parametric testing was initiated at 60 MW to assess the broad impacts of limestone injection, flue gas humidification, and sorbent recycle. Figure 3 shows average reductions achieved throughout the LIFAC process. About 22 percent SO₂ reduction is achieved in the boiler. This is increased to about 52 percent with humidification, and further raised to 75 percent with the use of sorbent recycle from the ESP ash hoppers. These tests were conducted with a fine grind limestone (80% minus 325 mesh) with a Ca content above 90 percent. A Ca/S molar ratio of 2.0 was held near constant and a 4 to 5° Fahrenheit approach to saturation was maintained in the activation reactor.

Figure 4 shows the impacts of varying the Ca/S molar ratio. The majority of the tests have been conducted at 2.0, but the trends are as expected. The higher the Ca/S ratio, the higher the SO₂ reduction. Results show, however, that SO₂ reductions of 75 to 85 percent are possible when spent sorbent is recycled and a 3 to 5° Fahrenheit approach to saturation temperature is maintained.

Figure 5 shows the impact of recycling spent sorbent under various boiler loads. The Ca/S molar ratio was maintained at about 2.0 and the level of humidification is high (4 to 5° Fahrenheit above saturation). Generally, there is an 18 to 25 percentage point increase in SO₂ reduction as a result of sorbent recycle. With recycle, total SO₂ reductions ranged from 75 to 85 percent depending on boiler load.

Although only preliminary testing has been completed, the results are encouraging. Additional work will be conducted to optimize these process parameters in hopes of maintaining a minimum of 80 percent SO₂ reduction at all boiler loads.

At this point it has also been shown at RP&L and other LIFAC installations that the system can be installed and operated without affecting normal powerplant operations. It will also be shown that the system can economically reduce SO₂ emissions when compared with other flue gas desulfurization technologies.

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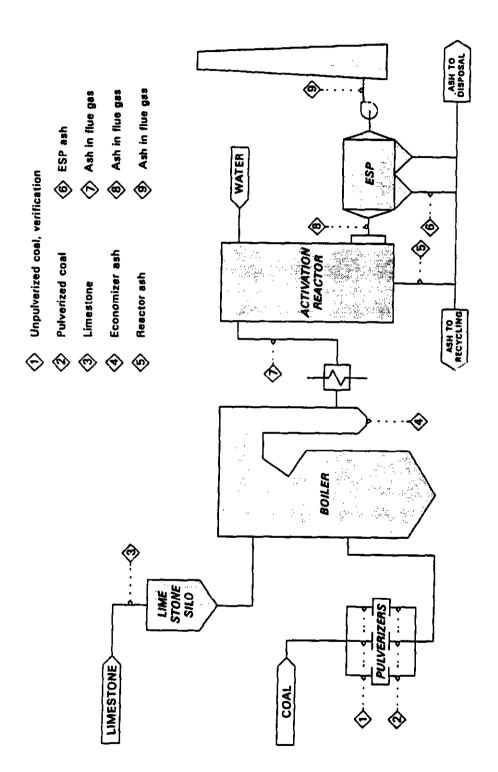


Figure 1. LIFAC RP&L DEMONSTRATION -- SOLIDS SAMPLING LOCATIONS

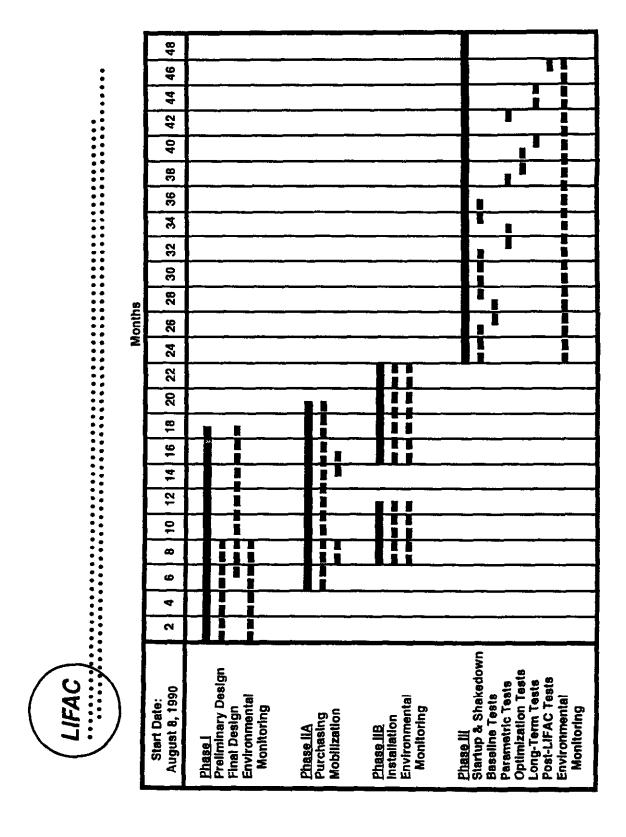


Figure 2. LIFAC DEMONSTRATION -- CURRENT PROJECT SCHEDULE

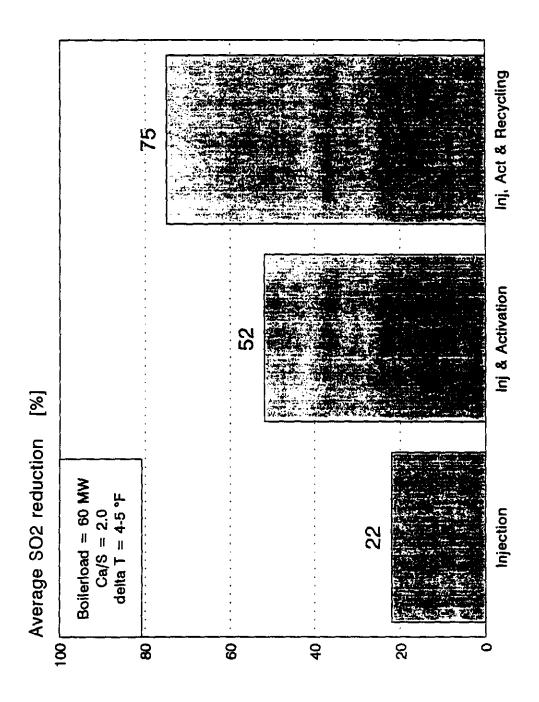


Figure 3. PARAMETRIC EFFECTS

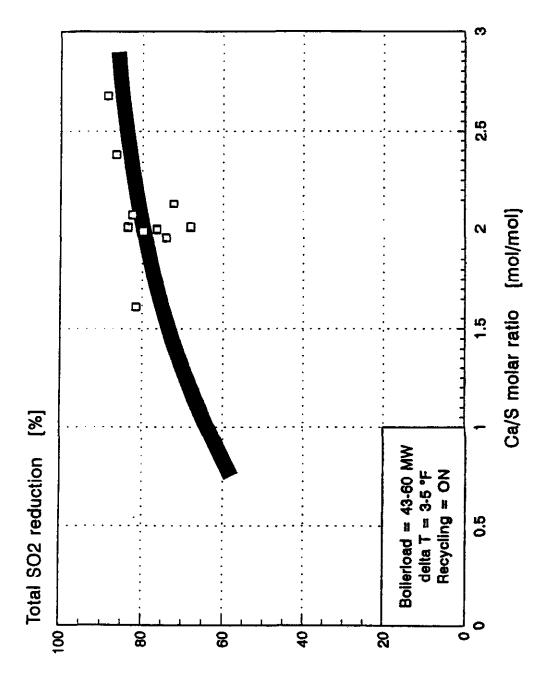


Figure 4. SO2 REDUCTION vs. Ca/S MOLAR RATIO

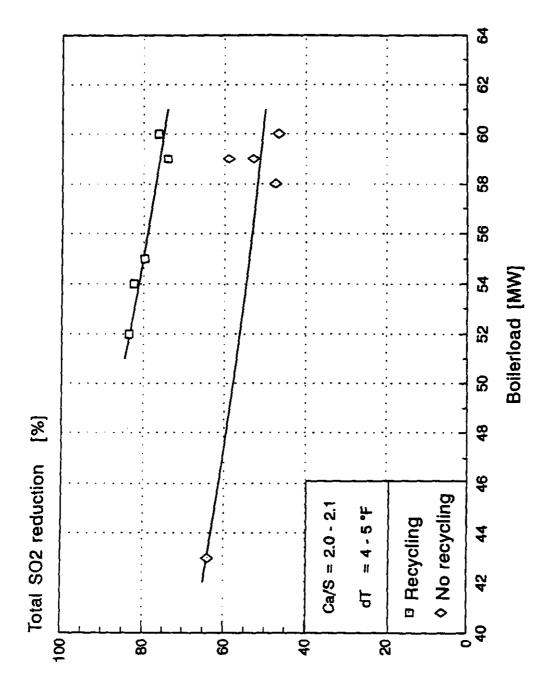


Figure 5. SO2 REDUCTION vs. BOILERLOAD



CHIYODA THOROUGHBRED 121 INNOVATIVE CLEAN COAL TECHNOLOGY DEMONSTRATION PROJECT INITIAL TESTING RESULTS

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ABSTRACT

The Chiyoda CT-121 Project at Georgia Power's Plant Yates Unit #1 is a Round II, \$36 million, Innovative Clean Coal Technology project co-funded by the Department of Energy, the Electric Power Research Institute and The Southern Company. The CT-121 scrubbing process features a single SO₂ absorption module, called the Jet Bubbling Reactor (JBR), made entirely of fiberglass reinforced plastics (FRP) where several chemical reactions (absorption/neutralization/oxidation/crystal growth) take place concurrently. The 100 MWe flue gas scrubber uses ground limestone to remove up to 95% of the SO₂ in the flue gas from a pulverized coal-fired boiler, producing a high quality gypsum by-product. Gypsum will be

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deposited in a gypsum "stack", a disposal technique commonly used in the phosphate fertilizer industry. Operational testing, continuing through early 1995, will include sustained high performance testing, simultaneous particulate removal in the JBR, alternate limestone and alternate higher sulfur coal. Initial results from parametric testing have demonstrated the excellent SO₂ removal and particulate removal performance of this unique flue gas desulfurization technology.

INTRODUCTION

As part of the Innovative Clean Coal Technology (ICCT) program, the Department of Energy (DOE), the Southern electric system, and the Electric Power Research Institute (EPRI) are sponsoring a 100 MWe demonstration of the Chiyoda Thoroughbred (CT-121) process. The \$36 million project is located at Georgia Power Company's Plant Yates Unit 1, near Newnan, Georgia.

This demonstration project began with the retrofit construction of a CT-121 wet-limestone scrubber to a 100 MWe pulverized coal-fired boiler. The CT-121 process involves the use of a unique process vessel called the Jet Bubbling Reactor (JBR) in place of the traditional spray tower/reaction tank arrangement of most conventional FGD processes. Start-up occurred in October, 1992 and the first phase of the demonstration began in January, 1993. The demonstration project is scheduled to continue through early 1995.

The demonstration is divided into two major periods with the first utilizing the pre-existing electrostatic precipitator (ESP), operating at full capacity. The second period, scheduled to begin in early 1994, will involve the field-by-field deenergization of the ESP and repetition of the tests executed in the first period. The second period will be used to evaluate the process' ability to remove uncontrolled particulate from a coal-fired boiler and the effect of high fly-ash concentrations in the slurry on scrubber performance. The two periods are further divided into parametric testing and long-term load-following testing. It is the results of the low-fly-ash parametric testing which are addressed in this paper (Performance, operability, and reliability evaluation are the focus of this demonstration, with performance characterization the specific focus of the parametric testing).

Thus far, the performance of the CT-121 process has been excellent, with SO₂ removal greater than 90% easily achievable and 98% SO₂ removal achieved under some operating conditions. Additionally, 90% of the particulate matter in the flue gas following the ESP was removed by the JBR. Long-term testing, currently in-progress, should help establish CT-121 as a highly reliable FGD process at a U.S. electric utility plant.

HISTORICAL OVERVIEW

The Yates Chiyoda project was one of four successful proposals from The Southern Company in Round II of the DOE's Clean Coal solicitation in 1988. Design and engineering began in 1989 at SCS and the Cooperative Agreement was executed in April of 1990. Construction began with ground breaking in August of 1990 and was completed in October of 1992 following a significant delay in obtaining permits from the State of Georgia's Environmental Protection Division. Operations began in late October of 1992 and continue today.

The interest in the Chiyoda process stems from The Southern Company's previous experience with five different FGD systems at Gulf Power's Plant Scholz in the late 1970's. The CT-121 process was selected because of its reliability and potential to offer significant cost reductions over other FGD processes. Lessons learned at Scholz and from other Chiyoda CT-121 FGD systems have been incorporated and expanded into the aggressive evaluation program now underway at Plant Yates.

FACILITY DESCRIPTION

The equipment comprising the demonstration facility can be divided into five major systems: boiler/ESP; CT-121 scrubber/wet chimney; limestone preparation circuit; byproduct gypsum stacking area; and process control system.

Plant Yates' Unit 1, with a rated capacity of 100 MWe, is the source of flue gas for the CT-121 process. All of the flue gas from this unit is treated by the CT-121 wet FGD process with no provision for flue gas bypass (The CT-121 process must remain in service whenever

the boiler is operating). During the low fly-ash phase of parametric testing, the existing ESP for Unit 1 is being used for particulate control. The design efficiency for this ESP is 98%.

A simplified process flow diagram for the CT-121 process is presented in Figure 1. The CT-121 employs a unique absorber design, called a Jet Bubbling Reactor (JBR), to combine conventional SO₂ absorption, sulfite oxidation, and gypsum crystallization in one reaction vessel. This significantly reduces the potential for gypsum scaling, a problem that frequently occurs in natural-oxidation FGD systems. Since much of the crystal attrition and secondary nucleation associated with the large centrifugal pumps in conventional FGD systems is also eliminated in the CT-121 design, large, easily dewatered gypsum crystals are produced.

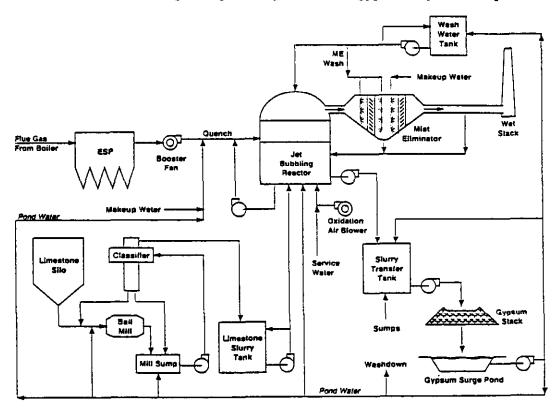


Figure 1. Simplified CT-121 Process Flow Diagram

In the Yates application, the flue gas enters the inlet gas cooling section after the I.D. fan. Here the flue gas is cooled and saturated with a mixture of pond water and JBR slurry. From the gas cooling section, the flue gas enters the JBR, the central feature of the CT-121 process. The gas enters the JBR through an enclosed plenum chamber formed by an upper

deck plate and a lower deck plate. Sparger tube openings in the lower deck plate force inlet flue gas beneath the slurry contained in the jet bubbling (froth) zone of the JBR vessel. After bubbling through the slurry, the gas flows upward through gas risers which pass through both the lower and upper deck plates. Entrained droplets in the gas disengage in a second plenum above the upper deck plate, and the cleaned gas passes to the mist eliminator.

After leaving the mist eliminator, the clean gas exits the system through a wet FRP chimney. Since the gas enters the chimney saturated with water, any heat loss in the chimney will result in gas cooling and condensation. Condensate in the chimney is collected by a system of aeordynamically designed internal "gutters" and is returned to the JBR.

A closed-circuit wet ball mill limestone preparation system is used to grind the limestone to a small enough particle size so that the amount of unreacted limestone needed in the JBR can be kept to a minimum. The baseline particle size criterion is 90% less than 200 mesh.

Slurry from the gypsum slurry transfer tank is diluted andpumped to a lined gypsum stacking area for dewatering and storage. The stacking technique involves filling a diked area with slurry for gravity sedimentation. The filled area is then partially excavated to increase the height of the containment dikes. The process of sedimentation, excavation, and perimeter dike formation continues on a regular basis during the active life of the stack. Process water is decanted, stored in a surge pond and returned to the process.

During normal operation of the FGD system, the amount of SO_2 removed from the flue gas is controlled by varying the JBR pressure drop (ΔP). The ΔP is adjusted by varying the JBR liquid level. Higher liquid levels result in increased SO_2 removal. The pH can also be varied to affect removal with higher pH's resulting in increased removal. Boiler unit load and flue gas SO_2 concentration also affect removal efficiency.

TECHNOLOGICAL ADVANTAGES

The CT-121 process, as constructed at Plant Yates, offers several technological and economic advantages over both conventional spray tower scrubbers as well as previously

constructed CT-121 systems. The innovations which provide these advantages include the use of fiberglass reinforced plastics, the unique JBR absorber, a novel wet FRP chimney design, and the use of a gypsum stack for byproduct dewatering and storage.

Fiberglass Reinforced Plastic

Many of the vessels comprising the Yates CT-121 installation are made of fiberglass reinforced plastics (FRP) to withstand the corrosiveness of the FGD process streams. Two of the vessels (the JBR and the limestone slurry storage tank) were constructed on-site in 1990 since their large size precluded roadway shipment. In a temporary facility, the contractor built the two vessels by layering materials on a slow moving "mandrel", alternating and mixing resin applications, fiberglass mat, fiberglass weave, fiberglass strand and chopped fiberglass in a carefully designed sequence over several week's time. FRP internals and wall penetrations (nozzles, man-ways, sample ports, etc.) were then installed by hand with additional FRP components to complete these homogeneous plastic vessels. Other FRP components that were small enough for shipment (two tanks, inlet duct section, chimney sections, etc.) were assembled and joined on-site at about the same time. Baseline data on the FRP vessels' acoustic emissions and a finite element analysis were taken for comparison to their future condition.

A distinct advantage of the FRP construction was that it eliminated the need for a flue gas prescrubber to remove chlorides, because the corrosion resistance properties of fiberglass are superior to those of alloys. This represented a large capital cost savings to the project.

JBR

The inherently high reliability of the CT-121 process eliminated the need for a spare absorber. This results in significantly reduced capital costs compared to spray tower systems which typically are built with 'spare' absorbers. Additionally, the JBR offers the distinct advantage of providing simultaneous SO₂ removal and particulate control. The JBR's high particulate removal efficiency may allow elimination of the ESP in new plant designs or make it a good choice for a retrofit to a plant with a marginally performing ESP.

Wet Chimney

The lower than normal mist loading inherent in the JBR design, combined with the unique design of the wet chimney, eliminated the need for flue gas reheat, another cost saving feature. Fluid dynamic modeling was performed to assist in the wet duct and chimney design. The resulting FRP wet chimney has a specially designed system of "gutters" which collect the liquid as it becomes un-entrained or disengaged from the flue gas and returns it to the JBR. This design effectively eliminates rain-out from the chimney.

Gypsum Stack

The FGD byproduct gypsum solids are disposed of by stacking. Stacking combines the advantages of ponding and landfills. Ponding has low operating costs and low capital equipment requirements, while landfills require less space and have less environmental impact. During the low fly ash parametric test period, handling, stackability, and trafficability of the gypsum stack were carefully monitored.

TEST OBJECTIVES

The parametric testing portion of the low fly ash test period has recently been completed.

The major objectives of the parametric test program were to:

- Correlate the effects of pH and ΔP , and evaluate the effects of boiler load on system performance;
- Correlate the effect of limestone grind on system performance;
- Monitor solids properties and gypsum stack operation;
- Evaluate particulate removal efficiency; and
- Demonstrate reliable operation of the CT-121 FGD system.

The test schedule was a full factorial matrix of the three primary test parameters (ΔP , pH, and load) which affect SO_2 removal efficiency. A full factorial matrix was designed to eliminate the need for a complex statistical analysis to evaluate the collected data. In addition to the full factorial matrix, the test plan also includes selected tests to evaluate limestone grind.

SCHEDULE/MILESTONES

After project selection in 1988 and execution of the Cooperative Agreement in 1990, construction of the Yates CT-121 took approximately 26 months. Operations began in October of 1992 with system shakedown, and parametric testing was conducted from January through March of 1993.

In March, long-term testing began (with the ESP fully energized) and will continue through the end of 1993. In early 1994, high-particulate operations (ESP deenergized) will begin and continue through early 1995. The high fly ash test phase will also contain both parametric and long-term testing periods. Extended monitoring of the groundwater and the gypsum stacking area will carry through 1996.

RESULTS

The parametric test period proceeded well, beginning on January 17, 1993 and ending on March 31, 1993. In general the process responded quickly and smoothly to load and process parameter changes, usually stabilizing within one hour of the change. The process' availability and reliability were both 98% due to a low equipment failure rate.

 SO_2 removal efficiency in the JBR is a function of JBR ΔP , pH, load, and inlet SO_2 . The parametric test results indicate that SO_2 removal increases with increasing pH and JBR deck ΔP , and that removal decreases with increasing load and inlet SO_2 concentration. Figure 2 illustrates the increase in SO_2 removal with increasing JBR ΔP as well as with pH. SO_2 removal however, does not increase with pH above a pH of 4.5 at the conditions tested. The figure shows that performance in excess of 90% SO_2 removal is easily achievable for the CT-121 process without the use of additives. Additionally, limestone utilization was consistently greater than 97% in the selected pH operating range (4.0 - 5.0) and exceeds that historically achieved in spray tower scrubbers. Oxidation, one of the keys to the excellent performance of this scrubber, was consistently 100%.

Although slurry pH and deck ΔP are the primary operational parameters for controlling SO_2 removal efficiency, inlet SO_2 concentration and boiler load (flue gas flow rate) have an effect

on scrubber performance. Figure 3 shows that SO_2 removal decreases with increasing boiler load at constant pH and ΔP . Because of decreased dynamic head (i.e., decreased pressure drop in the duct, plenum, etc.) at lower loads (due to lower gas flows), the JBR slurry level must increase to maintain a constant JBR ΔP at lower loads. Therefore, higher loads, which require a lower JBR level, result in decreased removal.

The SO₂ removal efficiency was seen to decrease with increasing inlet SO₂ at pH of 4.5. This trend is readily apparent in the regression analysis, but can also be seen in Figure 4 which compares tests at similar conditions with the exception of differing inlet SO₂ concentrations.

Particulate Testing Results

JBR particulate removal (ESP on) was evaluated in January, 1993, concurrent with parametric testing. The results indicate that the CT-121 process is an excellent particulate control device with removal efficiencies of 90% measured for those particles not collected by the existing ESP. Additionally, 99.9% of the particles greater than 10μ and 90% of those in the 1-10 μ range were collected. Based on this performance, it is anticipated that the high fly-ash operating period will also demonstrate excellent particulate removal characteristics as well.

Results Regression and Empirical Model

The parametric test series developed data on system performance over the entire range of expected operating conditions. These data were then used to construct an empirical model of system performance. Besides being a simple and efficient way of presenting the results of this test program, such a model is useful for choosing the most efficient operating conditions for a desired level of performance. It may also be possible to use this model to estimate emissions in situations where the CEM system has failed.

Linear models were investigated since linear is much simpler than non-linear regression analysis. The goal of this effort was to have a regression model in which all of the terms

were statistically significant and sensible from a technical perspective. The following equation was found to meet these criteria (Note that it is linear with respect to Number of Transfer Units (NTU) and becomes exponential when expressed in terms of efficiency):

$$SO_2$$
 removal (%) = $100*(1-exp-(A*pH + B*\Delta P + C*ISO_2 + D*LOAD + E*pH^2 + F*pH*\Delta P + G))$

Where: A,B,C,... = numeric coefficients

 ΔP = JBR differential pressure, in WC

 ISO_2 = Inlet SO_2 (@ 3% 02), ppm Load = Unit Gross Load, MWe

With an R^2 of 0.99 and no evidence of autocorrelated residuals, this model can easily be used to make accurate predictions of SO_2 removal performance over the range of operating conditions tested.

ECONOMIC EVALUATION

An important result from the parametric test series is the relationship between operating conditions, SO₂ removal performance, and operating costs. This information can be used to choose lowest cost operating conditions for a desired SO₂ removal, and for comparison with other system designs such as open spray towers. The analysis presented here is limited to limestone and power cost (e.g., I.D. fan, motors, pumps, ball mill) data. It does not consider O&M costs which are assumed to be constant.

Figure 5 shows that the cost per ton of SO_2 removed was relatively constant over the range of conditions tested with a standard deviation of 0.6. Most of the variability is the result of operation at different loads. At a given load, the fan power costs (normalized to \$/ton SO_2 removed) are seen to be relatively constant. Fan power does increase with both load and ΔP as expected; however, the increased SO2 removal at higher ΔP 's results in a relatively constant normalized power cost. A fan power credit, equivalent to the documented I.D. fan power consumption prior to the CT-121 scrubber construction, was taken when calculating power costs.

CONCLUSION

The Chiyoda CT-121 scrubber has exhibited excellent performance throughout the demonstration project, including the parametric test period. It easily exceeds 90% SO₂ removal, while maintaining consistently low variable operating costs and power consumption less than 1.5% unit Maximum Rated Capacity (MRC). Limestone utilization in excess of 97% is achievable at any pH within the established operating range of 4.0 to 5.0. The scrubber's ease of operation allows even the most inexperienced operator to quickly become familiar with system operations. Follow-on testing in a long-term load-following mode should firmly establish the viability of this unit as both an SO₂ reduction process, as well as a particulate control device. Initial indications are that the process' load following capabilities are excellent.

ACKNOWLEDGEMENTS

The authors wish to thank all those involved in the success of the Chiyoda CT-121 project at Plant Yates. Specifically, we are grateful for the tireless efforts and continued support of Messrs. Rob Moser and Jeff Stallings of EPRI, Mr. Hiroshi Yanagioka of Chiyoda-Japan, Messrs. Mike Knowles, Ross Kist, Steve Lee, Bill Mashburn, and Flip Kee (retired) of Georgia Power Company, Messrs. Randall Rush, Pete Honeycutt, Zach Looney, Lamar Larrimore, Kamyar Vahkshoorzadeh, and Steve Kimble of Southern Company Services, Inc., Messrs. Buddy Hargrove, Jack Burke, and Greg Stevens of Radian Corporation, and certainly the men and women of Georgia Power Company's Plant Yates who, on a daily basis, meet and overcome unfamiliar obstacles in support of scrubber operations.

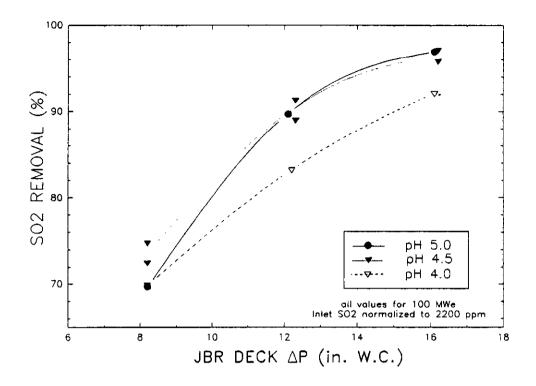


Figure 2. SO2 Removal vs JBR Deck ΔP

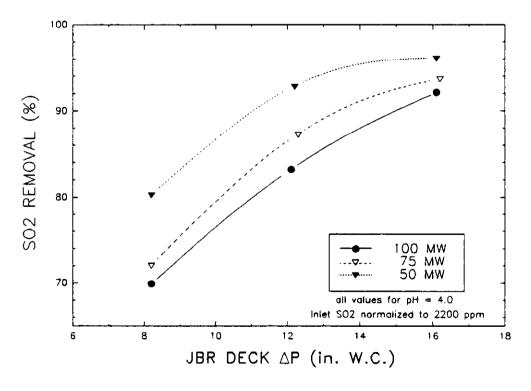


Figure 3. Effect of Load on SO2 Removal

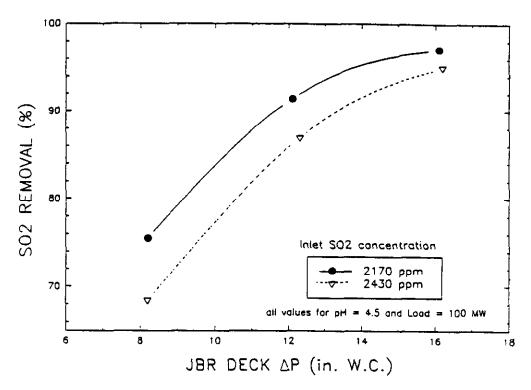


Figure 4. Effect of Inlet SO2 on SO2 Removal

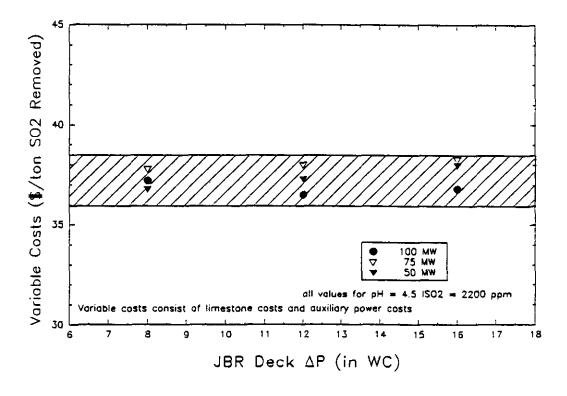


Figure 5. Variable Cost Deviation



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INDUSTRIAL POLLUTION CONTROL: 1993 PERFORMANCE UPDATE OF THE RECOVERY SCRUBBER

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ABSTRACT

A cement plant application of the Passamaquoddy Technology Recovery Scrubber¹⁸ pollution control process, an Innovative Clean Coal Technology Program project, began initial operation at the Dragon Products Company Inc. plant in Thomaston, Maine in December, 1990. During 1991 and 1992 several changes were made to improve on-line time and system reliability. Performance of the system, now in full time operation, is discussed. Results of flue gas scrubbing and waste reclamation are given. Changes that have been made, and their impact on system reliability are explained.

Marketing efforts and potential future applications are reviewed.

OVERVIEW

The Project

Information on the project goals, participants, location, cost, duration, and disposition is given in Appendix A, BASIC PROJECT INFORMATION.

The Technology

The Recovery Scrubber process was selected as part of Round 2, Innovative Clean Coal Technology Program. It is a wet flue gas desulfurization process that uses waste (fly ash, cement kiln dust, incinerator ash, biomass ash) as the chemical scrubbing reagent. Useful by-products that minimize or eliminate the need for landfill disposal of waste are produced by the scrubbing reaction. Tipping fees for consumption of waste produced by others, sale of useful by-products and emission credits, and "fee for service" pollution control, generally allow profitable operation of the scrubbing process.

DESCRIPTION OF THE TECHNOLOGY

General Information

Detailed description of the technology has been given elsewhere [see references at end of paper.] The following general information is provided as it relates to the current discussion.

These may include fly ash, waste cement kiln dust, incinerator ash, biomass ash from wood fired systems, and other similar wastes in solid or liquid form. Use of these wastes has the advantage of providing low cost reagent and income from tipping fees for consumption of waste. It also has the advantage of reducing, or in some cases eliminating, the volume of waste that must enter a landfill, thereby conserving valuable landfill space. Figure 1. illustrates basic process flows and system components.

Chemical Reactants

The alkali metals sodium or potassium, rather than the alkaline earth metals calcium or magnesium, are used for combination with sulfur from flue gas. Because calcium sulfate is not formed there is no gypsum scaling within the scrubber and no requirement for disposal of gypsum or scrubber sludge. Sodium or potassium form soluble compounds with recovered flue gas sulfur (sulfate) or hydrochloric acid. They will not cause scaling, and both potassium sulfate and potassium chloride are highly valued marketable by-products.

Solids Recovery

Calcium present in the waste will react to form calcium carbonate (limestone) by combining with carbon dioxide from the flue gas. This results in scrubbing of carbon dioxide from the flue gas. The product, essentially limestone, makes the spent reagent useful as raw material for use in cement manufacture or as starting material for manufactured aggregate for use in asphalt or concrete, thus eliminating the need to dispose of spent material in a landfill. Both the environmental advantage and the cost advantage of producing a useful by-product rather than a waste sludge are important.

Energy Recovery

Waste heat from the flue gas being scrubbed is recovered and used in the Recovery Scrubber process. Recovery of the waste heat allows for economical recovery of the soluble alkali sulfate salts by simple evaporation of solution and crystallization of dissolved solids.

Alkalis Recovered

Recovered alkali sulfate salts are removed from the process as solid salt crystals of potassium sulfate or sodium sulfate. In situations where chloride is present in the waste used as reagent, or in the flue gas being scrubbed, the product will include potassium chloride and/or sodium chloride, or diatomic chlorine may be produced for sale if desired. The various salts produced can be separated to enhance their resale value. All of these products have resale value. Potassium sulfate has the highest value at \$200-\$240 per ton wholesale or up to \$400 per ton retail.

Installation and Operation

The scrubbing process was installed with minimal impact on the operating cement plant. It is an "end of the pipeline" retrofit process. The only interconnect to the cement plant that might have curtailed operation is the physical tie in of the flue gas handling duct, however, the tie in was made during a routine kiln shut-down with no impact on kiln operation.

The Recovery Scrubber operates as an integrated unit, therefore, all subsystems in the process were operable at the outset with the exception of the crystalline product pelletizing equipment which was not necessary for operation.

The process control system is by computer with operator interface and ability to override as necessary. The control panel and display are located on the desk of the cement plant kiln operator for his use. No additional operator is necessary.

CHANGES MADE AND SYSTEM RELIABILITY

Changes made since initial start-up have been reported before. They include tray flatness, gas distribution, solids-liquid mixing, and tray washing. Additional changes made since the last report in this forum include mist elimination and fine tuning of gas distribution. These two changes have made the largest improvement in operation and are described below.

Mist Elimination

The initial mist elimination system was of the mesh pad type. It is an effective means of droplet removal from a gas stream. As arranged in the Recovery Scrubber, however, the mesh pad could not be effectively washed. Particulate collected from the gas stream accumulated on mesh pad surfaces and eventually obstructed gas flow. Frequent shut down for cleaning was necessary. The mesh pad was replaced with a chevron type mist eliminator that was configured so that it could be continuously washed with recirculated wash water. The wash water is periodically purged and replaced with clear water to prevent build-up of particulate within the circuit.

Operating time with the mesh pad was limited to one to two weeks and occasionally as little as four days. The system would be stopped, opened, allowed to clear flue gas from process areas, and manually cleaned during a six to twelve hour shut down period.

Since installation of the chevron mist eliminators there has been no stoppage because of mist eliminator operation. There have been a few brief stoppages to clear plugged nozzles in the mist eliminator wash water delivery system. These stoppages are minor, requiring only an hour to clean or replace nozzles, and are becoming much less frequent as debris is gradually purged from the pipelines carrying wash water. Operating periods between nozzle cleanings are now on the order of three months.

Fine Tuning of Gas Distribution

As noted in previous reports [see references at end of paper], baffles were

installed as a retrofit solution to inadequate gas distribution within the plenum under the tray reactor.

Initial design criteria called for differences in gas pressure not to exceed 0.1 inches of water at any point under the tray. The "as built" condition (which was not the "as designed" condition) exhibited pressure differences as large as one inch of water. Retrofit baffles were installed to redirect gas flow from areas of high pressure to areas of low pressure within the plenum. Distribution was corrected to yield differences of typically 0.25 inches or less, but with two corners where pressure remained low by as much as 0.5 inches. One brief shut down in May of 1993 was taken to install additional baffles as "fine tuning" of the gas distribution.

Prior to installation of the baffles the tray operated poorly. Flow of scrubbing slurry depends, in part, on the agitation provided by gas passing through the tray to keep slurry solids in suspension. Areas of low gas pressure provided little or no agitation of the slurry and sedimentation resulted. As tray surface was increasingly covered by sediment the operating pressure of the remaining tray increased to the point of having to stop for tray cleaning. Initially this period was a few days to a week. After installation of the baffles tray operation was markedly improved to periods of about a month. Now that "fine tuning" of the gas distribution has been accomplished the operating time exceeds three months and we are continuing. Additional adjustment may be necessary in the future if long term operation indicates any problems.

RELIABILITY

System reliability has improved markedly since initial start up and has changed by the largest measure since beginning operation in the spring of 1993. The project is designed for a thirty year lifetime so it is too early to give an in depth measure of reliability. Table 1, however, gives an indication of reliability as a function of percentage of time the scrubber is operating while the kiln is in operation, and as a percentage of waste cement kiln dust that no longer goes to landfill disposal.

TABLE 1
SYSTEM RELIABILITY
Performance Since Spring 1993 Start Up

Month %	On Line Time*	% of CKD Not Wasted**	Comments
April	65.0 % (100% cleaning time)	57.0	Scrubber did not start up until April 14th. Cleaning took place April 21 to 26.
May	78.6	85.5	Kiln down for kiln support repair. Cement plant raw material storage tank down, forcing scrubber off line to await repairs. Start fine tuning of baffles for gas flow distribution.
June	80.0	90.5	Finish fine tuning of baffles for gas flow distribution. Kiln down for trunion repair.
July 1-18th (to date)	95.9	96	-

^{*} Percentage of time both kiln and Recovery Scrubber are in operation. Scrubber may be off-line because of kiln operating conditions.

** Percent of CKD returned to the cement plant. This is all CKD not going to landfill disposal.

SCRUBBING AND WASTE RECLAMATION

Scrubbing

On line continuous monitors measure sulfur dioxide and nitrogen oxides on the inlet and outlet of the scrubbing system. Long term removal efficiency is 90 to 92 percent. If input sulfur dioxide concentration is below 50 ppm the indicated removal efficiency is below 90 percent. This is instrument inaccuracy rather than a real drop in scrubbing efficiency. For input levels above 100 ppm the observed removal is in the 92 to 95 percent range.

Nitrogen oxides are impacted by the scrubber to the extent of 5 to 15 percent removal. The removal is NO₁ rather than NO, and removal percentage changes as kiln burning conditions change.

Carbon dioxide is removed to the extent there is calcium sulfate or calcium oxide present that can be converted into calcium carbonate. For the cement plant this is 1 to 3 percent of the flue gas CO₁. For coal or oil fired boilers CO₁ removal would typically be in the 10 to 12 percent range.

Scrubber impact on volatile organics will be tested during August and September, 1993 by the U.S. EPA and separately by an independent laboratory for Passamaquoddy Technology L.P. Results can be reported during the Third Annual Clean Coal Technology Symposium.

Particulate emissions are very low. The methods used in this process for gas liquid contact and mist elimination both lend themselves to low particulate emissions. Stack tests for State compliance will be re-run in September, 1993 to verify current performance. Past testing showed emission levels below 0.006 grains/dscf. Current levels are expected to be lower by a factor of two or three, that is, 0.002 to 0.003. This compares very favorably with the both the current BIF regulation of 0.08 grains per dscf, and the proposed new standard of 0.015 grains/dscf.

Waste Reclamation

Reclamation of CKD, fly ash, and biomass ash are discussed in the following section. CKD is currently processed on a continuing basis. Fly ash will begin entering the system in August, and biomass ash in September or October. All of these wastes can be processed to provide benefit to both the cement plant and the waste generator.

Cement Kiln Dust

Cement kiln dust (CKD) is consumed at the rate it is produced by the cement plant, typically 100 to 250 tons per day. For CKD to be useful, and more importantly not detrimental, as raw material feed to the cement plant there are two primary requirements. First is that potassium (or in other plants potassium or sodium) present in the waste be removed so that it does not become part of the cement. This typically requires that potassium content in the waste be reduced to those levels found in normal raw material. It is permissible, however, for renovated CKD to have somewhat higher potassium levels because it usually constitutes a minor portion of the total feed. The second requirement is that sulfate levels in the waste be reduced before it is returned as raw feed. This is not an absolute requirement as sulfate is always added to cement during the

finish grinding process. Table 2 gives analyses for treated CKD for comparison with normal raw feed and raw feed composed, in part, of renovated CKD.

Table 2
Comparison of Renovated CKD, Type 1 Raw Material, and New Raw Material Containing Renovated CKD

		Type l Normal Raw Material	Type l Raw Material (90%) Combined With
Oxide	Reacted CKD	Typical	Processed CKD (10%)
SiO;	10.30	14.0	13.63
A1,0,	3.48	3.7	3.68
Fe ₂ O ₃	1.69	1.6	1.61
CaO	39.80	44.8	44.30
Mg0	2.84	3.0	2.98
SO ₃	4.38	0.3	0.71
K,0	2.21	1.1	1.21
Na ₂ O	0.37	0.4	0.39
Loss on Ignit.	33.61	35.0	34.86

Raw feed for a cement plant is made by inter-grinding a variety of raw materials in proportions that will yield a specified combination. As shown in Table 2 there are minor differences in feed prepared from treated CKD and normal raw feed. The differences, however, are small and are easily corrected by slight changes in the rate of addition of one or more of the mix components entering the raw material preparation process. Silica, for example is low by 0.37 percent. Increasing the rate of sand addition to the raw material grinding mill will correct the deficiency. Similarly limestone is low by 0.5 percent. Addition of limestone, in this case by 0.35 tons per hour in a 100 ton per hour system, will bring CaO into spec.

The result of CKD renovation is that feed prepared from processed material is entirely acceptable in cement manufacture.

Fly Ash

Fly ash from a coal and wood bark fired boiler will begin entering the system in August, 1993. The composition of the fly ash is very different from that required in cement manufacture. It is high in silica and alumina, low in calcium, and high in potassium. Use of the scrubbing process allows removal and recovery of the potassium without discharge to the environment. The fly ash, therefore, becomes a new silica source for the cement plant. Tipping fees received by the cement plant, based on the silica requirement for cement production, can be several million dollars per year. Savings for the fly ash generator are of similar magnitude, a win - win solution.

Biomass Ash

Biomass ash from a wood waste fired boiler will begin entering the system in September or October, 1993 if the current schedule holds. The ash currently costs more than \$50 per ton to dispose in a landfill, and continues to carry an unknown future liability. For the cement plant it will be a source of potassium for by-product production and a source of calcium, silica, and iron for addition to the cement plant raw material preparation system.

MARKET POTENTIAL

The market potential for this technology is quite large. Because the process will frequently operate at a profit it will, in many cases, be the lowest cost means of pollution control available. It is applicable to a variety of fossil fuel or waste fired facilities and can impact a number of industries including cement, power, paper, waste incineration, and heavy manufacturing. The most immediate market is likely to be the cement industry, although applications in pulp and paper and utility boilers are currently under evaluation.

As developers of Clean Coal Technology Projects are aware, marketing a new technology is a slow process. All of the concerns about new technology, reliability, energy costs, long term wear or corrosion, etc. apply. These concerns are compounded by the current state of the U.S. and World economy. There are no solutions to these concerns, except to be a proven and ready technology if and when industries are impacted by the need for pollution control or the high

cost of fuel.

Our efforts have centered on the U.S. and Canadian markets, but we have expended considerable effort in Europe and the new nations of Eastern Europe as well as in the Mid-East. We have provided detailed evaluations for, and visited most of, 31 industrial facilities where the Recovery Scrubber process is applicable. Our expectation is that these efforts will begin to bear fruit by year's end.

APPENDIX A

BASIC PROJECT INFORMATION

The Project Participants

The project participants are:

- The U.S. Department of Energy, Pittsburgh Energy Technology Center;
- Passamaquoddy Technology L.P., owner of the technology;
- Dragon Products Company Inc., a subsidiary of CDN U.S.A. and the host site providing partial funding of the installation.

Goals of the Project

Project goals were to design; build; operate and demonstrate the new Recovery Scrubber technology on a coal fired wet process cement manufacturing kiln; to eliminate landfilling of waste cement kiln dust, a waste product of cement manufacture; and to significantly reduce emission of flue gas sulfur dioxide from combustion of coal. Further goals are to assess the environmental and economic performance of the process.

Location

The project is located in Thomaston, Maine at the Dragon Products Company Inc. cement plant which is owned by CDN U.S.A. The area is a scenic Maine coastal town, heavily dependent on tourist trade and on remaining a scenic coastal community, where control of environmental pollution is of vital interest to both the State of Maine and local residents. The host plant is also located up wind from a Class 1 area in Acadia National Park and is regulated accordingly.

Project Cost

The project is currently in Phase III, the Operating Phase, and will continue in the Operating Phase for 2 months. Final project cost is, therefore, not yet available. The cost to date is approximately \$17 million. Total cost will exceed \$17 million when all project related costs associated with the operating period and final report are determined.

Project Duration

Construction began in April of 1990 (earthwork related to clearing the site began in the fall of 1989). The process was first operated nine months later on December 21, 1990. After system debugging and process modifications the operating period began on August 20, 1991 and will run for a period of 13 operating months. The operating period will include only that time during which the system is actually in operation. The cement plant has been shut down for several long, and several short, maintenance or inventory plant outages. Therefore completion of the operating period will require more than 13 consecutive calendar months.

Project Disposition

After completion of Phase III the project will continue to be operated by Dragon Products Company Inc. as the waste cement kiln dust and sulfur dioxide control system.

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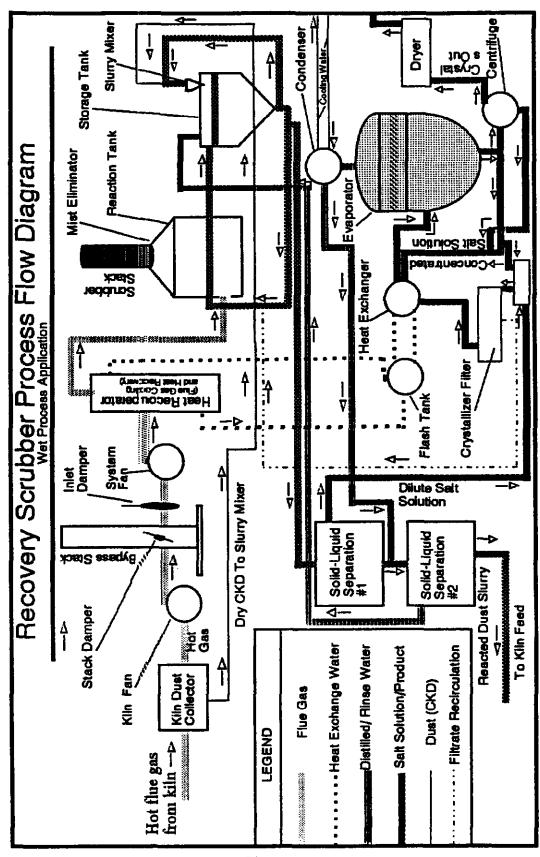


Figure 1

PAPER TO BE PRESENTED AT THE

SECOND ANNUAL CLEAN COAL TECHNOLOGY CONFERENCE

ATLANTA, GEORGIA September 8, 1993

Status of the Demonstration of Pulse Combustion In Steam Gasification

AUTHORS:

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ABSTRACT

ThermoChem's Clean Coal Technology project is a unique gasification process that uses indirect heating by combustion tubes immersed in a fluidized bed producing medium-Btu gas without needing an oxygen plant.

The concept of using pulse combustion tubes as an indirect heat source was developed by Manufacturing Technology Conversion International, Inc. (MTCI), who have licensed the technology to ThermoChem.

MTCI has completed a successful field testing of the pulse indirect heater (72-tube bundle) in a pulp and paper mill sludge/rejects gasification at Inland Container Corporation, Ontario, California in 1992. There is another field testing project of the pulsed indirect heater well underway in a distillery effluent treatment application aiming at zero-discharge by Esvin Tech, in Tamil Nadu, India. A third field testing of a three-heater (each with 72-tubes) fluid bed system for black liquor recovery is in the final stages of construction at a Weyerhaeuser paper mill in New Bern, North Carolina.

The proposed Clean Coal project is a scale-up of the pulse heater from 72-tubes to 252-tubes each. The Clean Coal gasifier would have 8 to 10 heater bundles to handle 300 T/D of dry coal.

Because of the large potential market for the ThermoChem process for the pulp and paper industry, the project was originally planned to the located in a Weyerhaeuser paper mill in Springfield, Oregon. After the project was selected under the Clean Coal Fourth round, ThermoChem requested DOE to move the project to the Caballo Rojo Coal mine site in Gillette, Wyoming to supply gas and steam

for "K-Fuel," coal-upgrading plant that would be built by Enserv, Inc., an affiliate of Wisconsin Power & Light.

The K-Fuel process upgrades low-rank coals producing a high Btu containing solid fuel called "K-Fuel" (to be substituted in power stations as low sulfur coal), and also generates wastewater and off-gas both of which need to be treated before discharge. The ThermoChem gasifier can not only use K-Fuel wastewater and off-gas, but it can gasify the fine coal that is not marketable or usable by the K-Fuel plant.

A preliminary test using K-Fuel effluent water and Caballo Rojo Coal fines was done in 1992 in MTCI's laboratory-scale gasifier facility in Santa Fe Springs, California at 20 lb/hr. This test showed that the organics in the K-Fuel effluent could be destroyed in the MTCI gasifier. Further testing in a larger facility (1,000 lb/hr) at Baltimore, Maryland is being planned for design verification of the process chemistry. A 252-tube bundle will be built and tested as part of the design verification in 1993.

PURPOSE OF TESTING

The purpose of the test run utilizing MTCI's gasifier facilities in Santa Fe Springs, California was to establish the following:

- 1. Efficient of the gasifier in destroying the organics found in the K-Fuels heat 2 water.
- 2. Produce gasifier char utilizing Caballo Rojo coal fines.

Detailed engineering of the demonstration facility will be completed by early 1994.

MTCI/THERMOCHEM BIOMASS STEAM REFORMING TECHNOLOGY

Manufacturing and Technology Conversion International, Inc. (MTCI) is an energy conversion and environmental control development company focusing upon the development of innovative technology applications based upon the phenomenon of pulsating combustion. Generally speaking, combustion instabilities are not only undesirable from both performance and environmental considerations, but can result in mechanical failures in the combustor or the furnace (boiler).

Over the years, many attempts have been made to harness those pulsations for a variety of applications. Many failed, a few were successful from the standpoint of performance but could not compete favorably in the marketplace. Some, primarily gas-fired home heating units, are available today but sales have been very sluggish in comparison to standard home heating systems.

About eight years ago, MTCI came to the realization that these combustion instabilities could provide many benefits when converted into well behaved oscillations. The company envisioned a host of applications for "stable" pulsating combustors; at first for clean and effective coal combustion, then for indirectly heated gasification systems and coal-fired fluid-bed combustors and finally for environmental control devices primarily aimed at coal-fired power plants.

In the following discussion, I will spend the first few minutes discussing pulse combustion and the performance and environmental benefits that can be derived therefrom. The rest of the discussion will be aimed at the specific applications available and finally to product improvements and development work now in progress.

PRINCIPLES AND BENEFITS OF PULSE COMBUSTION

The process of pulse combustion results from combustion-induced flow oscillations that are intentionally incorporated in combustor design to achieve process and system advantages for various combustion and gasification applications. The benefits accruing from controlled combustor oscillations are enhanced heat release rates (compact equipment), mass transfer rates (higher reaction rates, yields), heat transfer rates (indirectly fired heat exchangers), and the ability to develop a pressure boost that aids in reducing parasitic forced and induced draft fan power. The process has ancillary environmental benefits in drying applications, ash agglomeration, enhanced sulfur capture by dry sorbents, soot blowing and filter/baghouse cleaning.

The pulse combustor type used by the MTCI and ThermoChem equipment design is based on the Helmholtz configuration (<u>Figure 1</u>). The basic configuration consists of an aerodynamic air inlet valve (fluidic diode), a combustion chamber, and a tailpipe (or resonance tube). The combustion chamber and the resonance tube comprise a Helmholtz enclosure having a quarter-wave resonant frequency. There are no moving parts (flapper valves) thereby making it ideal for coal combustion as well as for other solid, gaseous and liquid fuels. The selection of this configuration was made primarily because of its excellent suitability and reliability for coal burning.

In conventional coal burners (cyclone, vortex, bluff body, etc.) combustion efficiency is highly dependent on the flow pattern and the extent of the relative motion between the burning coal particle and the surrounding gases. As the coal particles burn, they become smaller and increasingly ash-laden (char) while oxygen concentrations are decreasing. Oxygen diffusion from the surrounding gas to the burning ash-laden char particles also decreases requiring additional residence time and turbulence to achieve higher carbon burnout. This is caused by a boundary layer of products of combustion ($\mathrm{CO_2}$ and CO) forming a diffusion barrier between the oxygen and the smaller ash-laden coal particle. The entrainment prone nature of small particles, as carbon depletes from the burning coal particle, prevents significant relative motion between the particle and the surrounding gases, requiring the expenditure of high levels of parasitic power to create the flow patterns and forces necessary to drive the combustion process to completion.

In pulse combustion, the oscillating flow field, itself, provides high oscillatory relative motion between the burning coal particles and the surrounding gases. The boundary layer formed by the products of combustion, leaving the burning particle, is quickly swept away leaving little to no diffusion barrier as an impediment for oxygen reaching the burning coal particle. The reaction rate is, therefore, essentially kinetically limited rather than diffusion limited. Heat release rates can reach as high as 6 MMBtu/hr.cu.ft., more than an order of magnitude higher than in conventional combustion processes. This renders pulse combustors very compact and lower in capital cost. Combustion of

standard grind pulverized coal has been achieved in 30 to 40 ms. In conventional coal burners, residence times in the order of \(\frac{1}{3} \) to 1\(\frac{1}{3} \) seconds are required.

In conventional combustor and fire tubes arrangements, essentially all the heat is released by burning the fuel in the combustor. The heat is stored in the form of sensible heat in the flue gas which is at its peak temperature at the inlet to the fire tubes. This requires the use of a high-temperature material at the inlet region of the fire tube. As the heat is transferred from the flue gas through the fire tubes, the temperature of the flue gas monotonically decreases along the length of the tube. In this case most of the heat transfer on the flue gas side of the tube is convective. Radiant heat transfer may take place near the fire tube inlet if the gas is hot enough to be significantly radiant. In pulse combustion, however, not all the fuel burns in the combustion chamber but combustion persists down the resonance tubes (fire tubes) for a significant length in an oscillating flow field environment. Thus, for the same heat transfer duty, the inlet flue gas temperature to the resonance tubes is lower than in the case of conventional fire-tube systems, but the continued heat release from burning fuel in the resonance tubes maintains a higher bulk flue gas temperature than in the conventional case. Radiant heat transfer will also maintain to a longer length on the flue gas side of the resonance tube. In addition to the enhanced radiant heat transfer component along the resonance tube, a large enhancement in the convective heat transfer component is also achieved due to the oscillatory flow field of the gases. The enhancement in connective heat transfer results from an increase in both the average velocity (caused by the combustion-induced pressure boost), and the superimposed oscillatory velocity component (scrubbing of the boundary layer).

<u>Figure 2</u> represents experimental heat transfer data obtained on a gasifier combustor heat exchanger. The figure represents a comparison of experimental data with theoretical non-pulsating flow values. Actual enhancement of the heat transfer coefficient was about 3 to 5 times higher than that achieved by similar indirectly heated systems.

An important benefit of enhanced heat transfer rate is the ability of the reactor to support highly endothermic reactions such as the carbon-steam reaction. Rapid heat transfer to the fluidized bed material being processed results in very high rates of devolatilization and pyrolysis. This, in turn, results in the formation of char particles that are extremely porous with high reactivity. Steam reacts with the char to provide a synthesis gas mixture containing $\rm H_2$ and $\rm CO$. Devolatilization and gasification reactions are highly endothermic reactions. High heat transfer rates are therefore essential to support such endothermic reactions in an economically viable reactor with a reasonable throughput.

Pulse coal combustors, properly designed, have been established to be low NO generators. NO levels as low as 83 ppm (@ 3% O_2 in the flue) have been achieved by MTCI in pulse combustion of coal and in the 10-25 range when fired with natural or synthetic gases. There are a number of combustion process related characteristics of pulse combustion that are relevant to NO production. The rate of combustion in these devices is sufficiently high, with short residence times, such that NO formation is reduced. NO formation is endothermic with limited kinetic rates and hence the shorter the residence time, the less NO formation during the combustion process. The pulse combustion process inherently contains both flue gas recirculation and reburn characteristics. During a portion of the cycle of the pulse combustor, flue gas returns to the combustion

chamber from the resonance tube mixing with the fuel and air prior to ignition by the hot combustion chamber inner surfaces to trigger the next portion of the combustion cycle.

The equivalent of reburn is caused by the burning of particles after they leave the combustion chamber. Measurements of temperature profiles along the combustor length suggested that 15 to 25 percent of the heat release takes place in the tailpipe. The flow environment in the tailpipe is also oscillatory providing an intense mixing during the reburn portion of the process, leading to further reductions in NO_x formed from both fuel-bound nitrogen and thermal sources in the combustion chamber. Figure 3 gives the NO_x levels obtained in the 72-tube pulse combustor.

PULSE COMBUSTION APPLICATIONS

The following discussion addresses the hardware and technology applications based upon the essential principles of pulsed coal combustion. A summary of the related MTCI pulse combustion-based technology is provided in $\underline{\text{Table 1}}$. For each application cited, process data and/or hardware has been successfully acquired and operated. The presentation is intended to provide a perspective that relates to the available technology data base and equipment maturity.

Indirectly Heated Thermochemical Reactor and Processes

This technology is comprised of a fluid-bed reactor that is indirectly heated by a heat exchanger that is comprised of the multiple resonance tubes of a pulsating combustor as shown in <u>Figures 4</u> and $\underline{5}$. In this design the multiple pulse combustor resonance tube heat exchanger is fired with a portion of the product gas produced in the fluid-bed reactor or other fuel available. The module has multiple aerodynamic valves.

The reactor is employed for a number of patented endothermic processes that are also listed in the table. The status of the technology is as follows. A commercially configured, full-scale heater module (5-8 MMBtu/hr) powering a 12-tons/day fluid-bed reactor (40 $\rm ft^2$) has been built, tested and demonstrated at the MTCI facility in Santa Fe Springs, California (Figure 6). This is a pilot unit that can be used at the facility for feedstock characterization, yield optimization and other system parameter information.

A smaller process development unit, 30-100 lbs/hr is also available at the Santa Fe Springs facility. This unit is primarily used for initial process development and characterization (all input and output streams).

A 17 ton/day gasification unit has been installed at the Inland Container Corporation facility at Ontario, California. This unit has been in operation since March 1992 and a long-term system test was conducted in July 1992. The system processes an industrial recycle paper mill sludge containing 50 percent solids, fiber rejects with plastic and old corrugated container lights (OCC). A photo of the system in operation is provided in Figure 7. Tables 2 - 5 present the operating parameters for a 500-hour test on this unit. This unit was modified to process black liquor and was tested at Inland with liquor trucked from the Simpson-Samoa mill. After these successful field tests, this heater development unit was moved to MTCI's Baltimore, Maryland facility. NREL-

sponsored straw or grass and woody biomass gasification tests and NSSC sulfite liquor tests for MEAD Container Board are planned for October 1993.

In addition, a 50 ton/day expandable to approximately 100 tons/day with the addition of two additional heat exchanger modules is being assembled at Weyerhaeuser's paper mill in New Bern, North Carolina (Figure 8). This unit processes black liquor from the pulping process, recovering energy from the lignin in the spent pulping liquor as well as process chemicals (sulfur and sodium) for reuse in the pulping process. A similar unit is now in operation for a bagasse-based spent liquor recovery process at an SPB pulp mill in Erode, Tamilnadu, India.

For coal gasification, ThermoChem, an MTCI licensee of the gasification technology, has been selected to negotiate a Clean Coal IV Demonstration Project utilizing the MTCI indirectly heated gasifier. The cost of the project, \$42,000,000, will be provided by the U.S. Department of Energy (\$18,700,000) and Ensery (\$23,300,000). Ensery is a subsidiary of the Wisconsin Power and Light Company. The gasifier, sized at approximately 300 dry tons/day of subbituminous coal will be located at the Caballo Rojo coal mine in Gillette, Wyoming and is intended to provide a product gas for electricity generation from boilers with the waste heat from the gasifier producing a high pressure (1150 psi) steam for a coal beneficiation process. The low-cost hydrocarbon-laden wastewater from the beneficiation process will also be processed in the gasifier as a source of steam for the reaction permitting recovery of the energy and sensible heat and destruction of organic toxics. An overall material and energy balance for the process is provided in <u>Table 6</u>. A simple schematic of the gasifier is shown in Figure 9. The tube exchanger bundles to the reactor contain over 250 tubes each for providing the endothermic heat of reaction.

The versatility of the MTCI Thermochemical reactor/gasifier for processing a wide spectrum of carbonaceous materials can be derived from <u>Tables 7</u> and <u>8</u>. A generalized schematic of the process is shown in <u>Figure 10</u>. Table 3 provides test data from lignite, subbituminous coal (Black Thunder, BT) and char as well as for a mild gasification process designed to provide a suite of gaseous, liquid and solid fuel products. Table 4 provides data for a variety of biomass and waste materials including Refuse Derived Fuel (RDF) and municipal wastewater sludge. <u>Table 9</u> indicates the levels of dioxin and furan reductions achieved in the gasification of chlorine biomass wastes. The tests were conducted with a paper mill waste sludge feedstock.

<u>Figure 11</u> shows integration of the ThermoChem gasifier with the K-Fuels process.

A preliminary test using K-Fuel effluent water and Caballo Rojo coal fines was done in 1992 in MTCI's laboratory-scale gasifier facility in Santa Fe Springs, California at 20 lb/hr (Figure 12). This test showed that the organics in the K-Fuel effluent could be destroyed in the MTCI gasifier. Further testing in a larger facility (1,000 lb/hr) at Baltimore, Maryland is being planned for design verification of the process chemistry. A 252-tube bundle will be built and tested as a part of the design verification in 1993.

Purpose of Testing

The purpose of the test run utilizing MTCI's gasifier facilities in Santa Fe Springs was to establish the following:

- 1. Efficiency of the gasifier in destroying the organics found in the K-Fuels heat 2 water.
- 2. Produce gasifier char utilizing Caballo Rojo coal fines.

Test Facilities and Feedstock Coal

The test facilities included a steam generator, a gasifier vessel with ThermoChem's single-tube pulse combustor, cyclones for char collection, and a venturi scrubber for condensation of water vapor (see Figure 12).

Summary of Analytical Results

- 1. Based on the leaching tests, none of the chars would be considered hazardous waste by EPA.
- 2. The compositional analysis indicates small quantities of aromatic hydrocarbons (intermediate products of the coal gasification), and inorganic constituents normally found in coal ash.
- Leaching tests indicate the organics found in the char are not readily leached out and the inorganics are typical of alkaline coal ash leachates.
- 4. Although not specifically tested, the carbon content and fineness of some of the chars would warrant design consideration to manage the dustiness and reactivity with oxygen prior to disposal.

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TABLE 1:

SUMMARY OF MTCI PULSE COMBUSTION-BASED TECHNOLOGIES AND APPLICATIONS

TECHNOLOGY	DESCRIPTION	APPLICATIONS
Indirectly heated thermochemical reactor	Multiple resonance tube gas-fired pulse combustor heating a fluid-bed thermochemical reactor	 Biomass steam reforming Low-rank coal steam reforming/gasification Black liquor recovery (Pulp & Paper) Mild coal gasification Catalytic steam reforming of heavy end residual hydrocarbons Sewage sludge steam gasification Industrial sludge processing Indirect drying Toxic waste to energy processing Steam gasification of RDF
Pulsed Atmospheric Fluid Bed Combustor (PAFBC)	A hybrid combustion system employing a pulse coal combustor and a fluid-bed combustor	 Clean combustion of low-quality crushed coal fuels
Tandem slagging pulse coal combustor	Two pulse combustors that operate in the slagging mode for ash rejection. The combustor operates out of phase to cancel pressure oscillations emanating from the tail-pipes in a decoupler/slag chamber	Industrial, oil and gas designed boiler, retrofit for clean coal firing
Multiple-resonance tube coal-fired pulse combustors	Pulse coal combustor having one or multiple aerovalves and multiple resonance tubes	 Commercial boiler retrofit applications Indirect-fired gas turbine

TABLE 2: Summary of the Overall System Operation for Test 500-hour Test (July 1992)

TOTAL HOURS FOR PULSE COMBUSTOR OPERATION:	516 Hours
TOTAL HOURS FOR SLUDGE FEEDING:	432 Hours
TOTAL WEIGHT OF SLUDGE FED:	275,730 Pounds
AVERAGE SLUDGE FEED RATE:	640 lbs/hr

TABLE 3: Typical Material Flow Summary for 500-Hour Test (July 1992)

	(JULY 1992)	
INPUT	lbs/hr	MMBtu/hr
SLUDGE FED	5 00 - 9 00	2.30 - 3.2
FEED MOISTURE (% wt.)	50% to 75%	
STEAM FOR FLUIDIZATION	1700	1.94
NATURAL GAS TO PC	350 - 360	7.5 - 7.7
(based on LHV)		
OUTPUT		
PRODUCT GAS	367 - 700	3.1 - 5.8
STEAM	4000	5.0
LOSSES		1.0 - 2.0

TABLE 4: TYPICAL PRODUCT GAS ANALYSIS

	(JULY 1992)	
AVERAGE BED TEMP.	(°F) 1515	1470
GAS COMPOSITION	(%V)	(%V)
H ₂	34.7	44.3
CĤ,	11.6	5.4
CH ₄ CO	22.5	18.1
CO ₂	27.0	29.8
C ₂ [*]	4.3	2.5

TABLE 5: PULSE COMBUSTOR DATA

			<u> </u>	
	(J	ULY 1992)	
FIRING RATE (HHV)	=	8.20 -	8.45 MMBtu/h	ır
(LHV)	=	7.4 -	7.7	
FREQUENCY	=	62 Hz		
PEAK-TO-PEAK	=	4 psi		
FLUE GAS EMISSION,	DRY	BASIS		
<u>Conditions</u>		<u>#1</u>	<u>#2</u>	<u>#3</u> 0.3
0 ₂ (%v/v)		1.4	1.8	0.3
CĎ (ppm)		23	0	97
NO _x (ppm @ 3%	0_{2})	25	30	32
$SO_2^{}$ (ppm)	-	0.0	0.0	0.0

TABLE 6:

OVERALL MATERIAL AND ENERGY BALANCE
FOR STEAM REFORMING OF SUBBITUMINOUS COAL

	MASS	ENTHALPY	HHV	
INPUT	(lb/hr)	(KBtu/hr)	(KBtu/hr)	
Coal	35,714	300,000	300,000	
Process Water	52,191	31,943	6,741	
Boiler Feedwater	73,929	15,007		
Vent Gases	5,582	16,486	15,094	
Combustion Air	127,044	0		
TOTAL IN	294,460	363,436	321,835	
OUTPUT				
Product Gas	31,250	188,352	187,834	
Steam @500 psi	33,202	41,466		
Steam @ 1150 psi	49,726	64,296		
Sulfur	332	1,322	1,322	
Char/Ash	2,817	16,958	16,095	
Solids from Scrubber	232	1,742	1,738	
Water from Venturi Scrubber	17,489	739		
Condensate from H2S Removal	1,450	48		
Flue Gas to Stack	157,916	17,766		
Heat Rejected in Cooler		24,117		
Heat Losses		6,630		
TOTAL OUT	294,414	363,436	206,989	
CLOSURE, percent	100.0	100.0		
Cold Gas Efficiency	57.6	% (HHV of Gas	s - HHV of Vent (Gas)/ HHV of Coal
Overall Thermal Efficiency	80.9	%		

TABLE 7:

ANALYSIS FOR FEEDSTOCKS TESTED IN PULSE-ENHANCED INDIRECT GASIFIER

Feed Material Lignite Lignite Lignite Sand Temperature (f) 1370 1430 7.3 1430 15.1 7.3 1430 15.1 7.3 1430 15.1 7.3 1430 15.1 7.3 145.1 7		BT Coal Dir. Gasif Limestone 1370 16.9 16.9 28.3 1.67 85.9 23.0	Char Linestone		BT Cost Mild Gasif
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Rate 15.0 15.1 15.0 15.1		18.90 16.9 16.9 1.67 1.67 23.0	Limestone 1254		
Rate		16.9 28.3 1.67 88.9 23.0			reto.
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## Rate (1b/hr) 30.6 ## Ratio		28.3 1.67 85.9 23.0	0.42		0.06
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siffcation Eff. (%) 96.1 44.4 4		23.0	2.21		
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1 tane 0.01 0.01 0.01 0.00 0.00 0.00 0.00 0.0		0.32	00.0	: -	7 75
10.04 10.01 10.03 10.03 10.00 10.00 10.00 10.00 10.00 10.00 10.00	0.0	0.15	0.0	0.0	7
tane 6.00 8.00 me 0.00 0.00 0.00 0.00 0.00 0.00 0.0	90.0	0.05	0.00	0.00	70.1
ane ane D.03 D.00 0.00 0.00 0.00 0.00	0.00	0.00	0.00	0.00	0.38
74 0.00 0.00 0.00 0.00 0.00 0.00 0.00 0.	0.00	10.0	0.00	0.00	00.0
9.00 0.00 0.00 0.00 0.00	20.0	0.103	0.00	0.0	0.69
0.00 0.00 0.00 0.16	9.6	0.00	0.00	0.00	0.32
0.00 0.00 0.16	3.5	5,0	0.00	9.6	0.64
ле 0.16 0.16		3,00	0.0	90.a	90.0
0.16		5.6	8.8	9.0	9.0
20 001	0,26	0,16			. c
00.001	100.00	100.00	100.00	100.00	100.00
Ory Gas (SCFM)	***		,		
279.2	, L	0.0	16.6	7.1.5	7.7
Carbon (1b/hr) 5.6 3.1	3.1	7.607	14.7	4.4.5	5.0.5
		:		•	٧.

* Nitrogen was used as fluidizing gas for mild gasification for char production

TABLE 8:

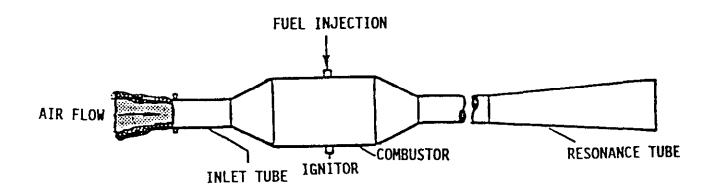
PRODUCT GAS COMPOSITIONS AND YIELDS FOR BIOMASS TESTS CONDUCTED IN THE MTCI INDIRECTLY HEATED GASIFIER

					RECYCLE	RECYCLED	KRAFT	RDF	HSM	MSM
COMPOSITION	PISTACHIO	P I S T A C H I O	000A	RICE	MILL FIBER	WASTE PAPER	HILL	SAND	SAND	LIHESTONE
(Vol.X)			CHIPS	HULLS	WASTE	W/PLASTIC	St UDGE	BEO	BED	BED
	37.86	35.04	48.11	42.83	38.86	50.50	52.94	45.54	55.21	54.40
	18.84	23.43	22.91	19.67	23.34	19.26	11.77	25.26	28.10	25.46
	28.73	25.20	20.18	24.40	23.27	20.10	21.94	14.51	5.95	5.66
	10.65	11.31	8.32	11.56	8.31	8.42	8.95	8.30	5.00	5.86
	3.92	5.02	0.48	1.54	6.40	1.72	3.00	6.38	5.74	8.62
TOTAL	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
	370	406	329	367	412	364	372	418	374	448
TEMP. (F)	1317	1216	1286	1326	1250	1326	1250	1450	1410	1306
YIELD	94.1	92.1	93.0	N/A	86.8	N/A	56.0	83.6	93.7	83.8
(X Carbon)										

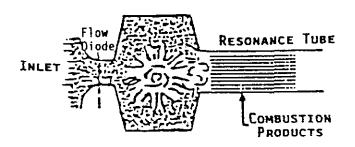
TABLE 9:

FURAN/DIOXIN LEVELS IN SLUDGE FEED AND GASIFIER EFFLUENTS

0000	670	7.2	9.7	0.33
HPCDO	150	2.9	9.5	N/D
HXCDD	280	N/D	14	Q/N
PC00	69	N/D	4	N/D
2,3,7,8 TCDD	33	N/D	27	0.07
ICDD	74	N/D	53	0.23
TOTAL DIOXIN	1543	10.1	100.2	0.33
OCDF	54	N/D	N/0	0.03
HPCDE	M/D	N/D	N/D	N/D
HXCOF	N/D	1.1	N/D	0.05
PCDF	110	N/D	7.0	N/D
2,3,7,8- TCDF	84	N/D	4.0	N/D
TCDF	440	N/D	170	0.46
TOTAL FURAN	550	1.1	171	0.51
	FEED SLUDGE	BED MATERIAL 1.1	CYCLONE ASH	CONDEN- SATE



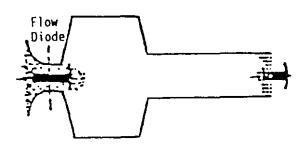
COMBUSTION CHAMBER

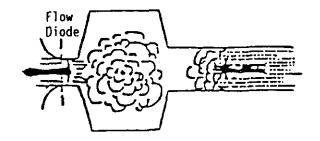


BACKFLOW

A. IGNITION AND COMBUSTION







C. PURGE AND RECHARGE

D. RECHARGE AND COMPRESS

FIGURE 1: PULSE COMBUSTOR CONFIGURATION

• - TUBE TO AMBIENT AIR

- TUBE TO PET. COKE IN STEAM

× - PREDICTED FROM QUASI-STEADY

STATE MODEL

TUBE TO WATER

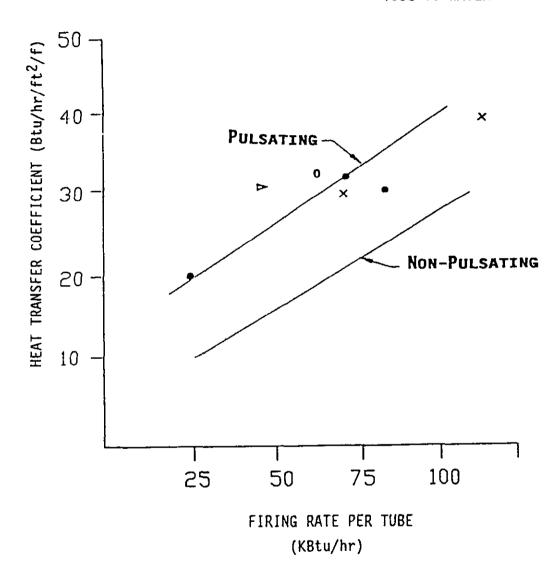


FIGURE 2: THEORETICAL HEAT TRANSFER COEFFICIENTS
UNDER NON-PULSATING CONDITIONS COMPARED
TO MEASURED DATA IN PULSATING FIRE TUBES

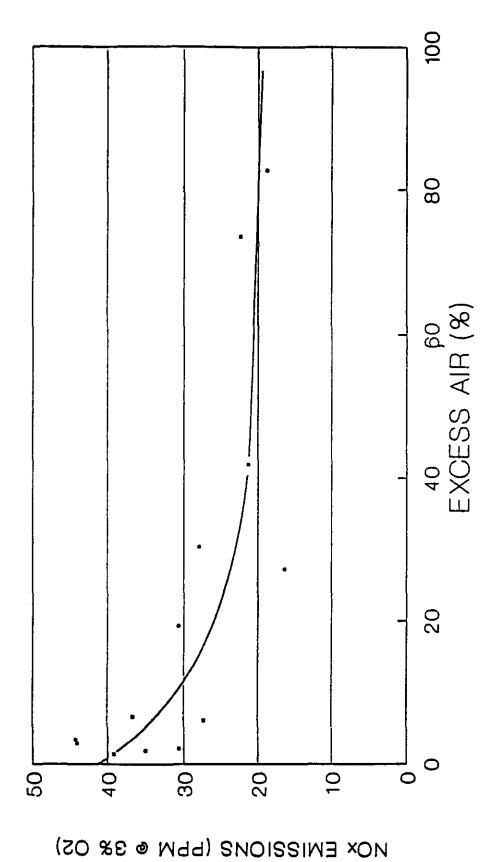


FIGURE 3: EFFECT OF STOICHIOMETRY ON NOX EMISSIONS FROM A PULSE COMBUSTOR

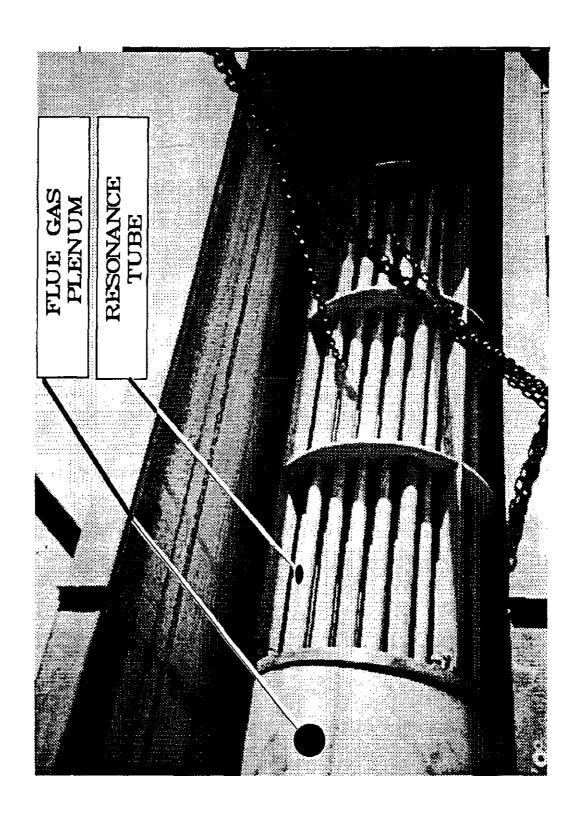


FIGURE 4: MULTIPLE RESONANCE TUBE PULSE COMBUSTOR

FIGURE 5: GASIFIER HEATER MODULE

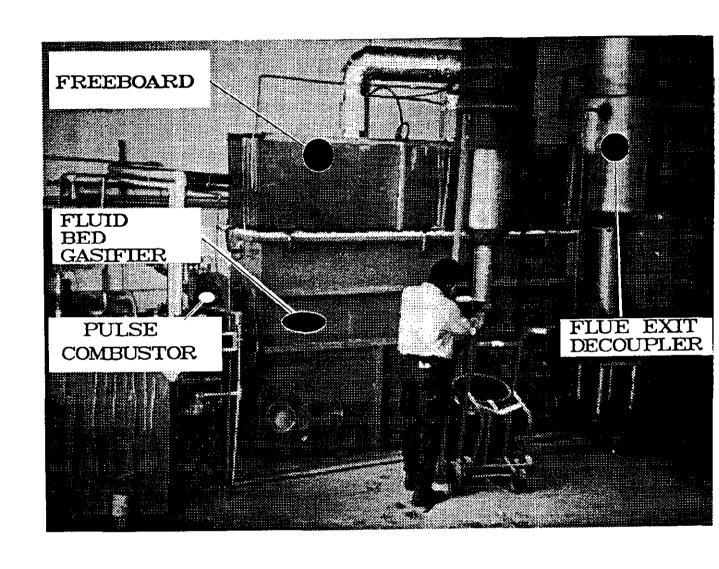


FIGURE 6: INDIRECTLY HEATED GASIFIER PILOT UNIT (12 TONS DAY)

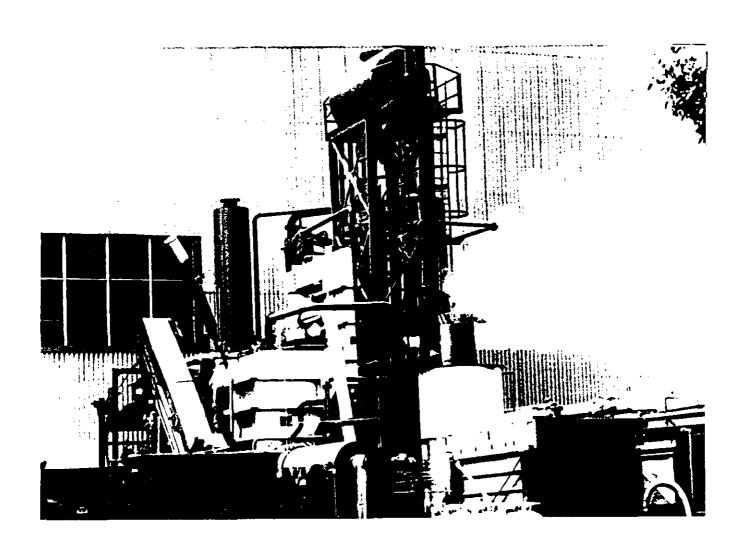


FIGURE 7: INLAND CONTAINER CORPORATION GASIFICATION UNIT (24 tons/day gasifier)

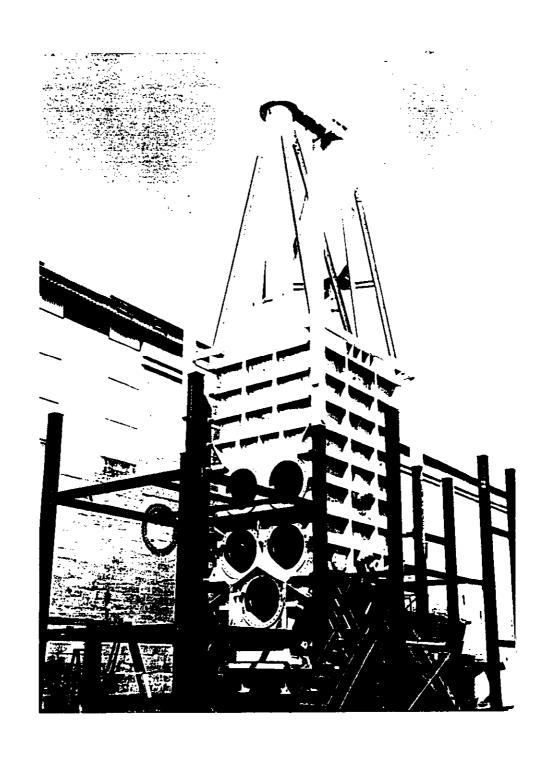


FIGURE 8: BLACK LIQUOR UNIT FOR WEYERHAEUSER

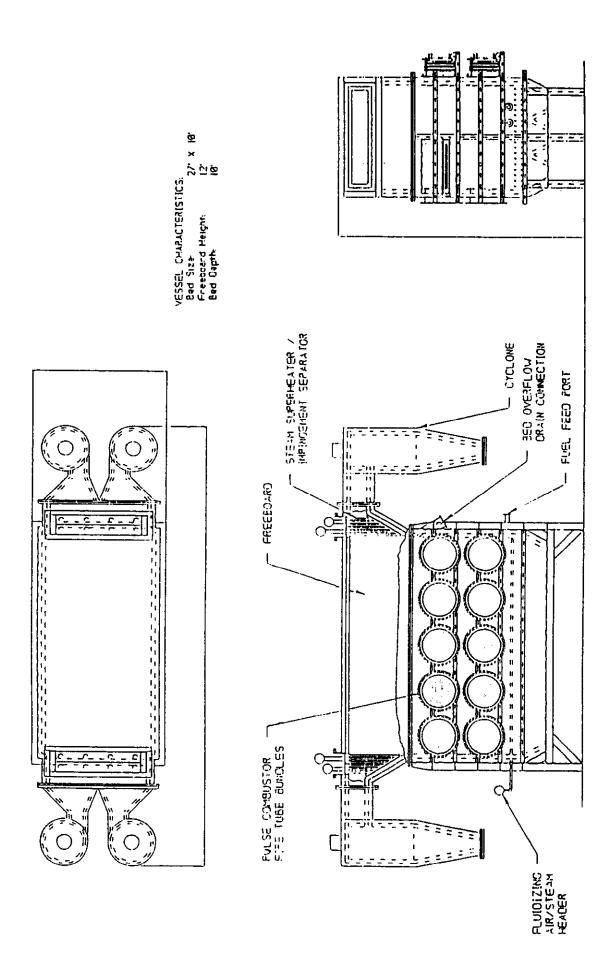


FIGURE 9: 300-TON PER DAY GASIFIER

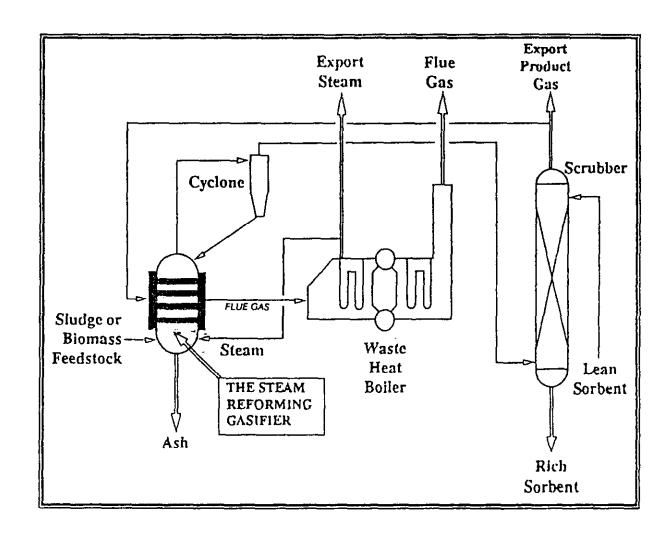


FIGURE 10: SIMPLIFIED SYSTEM SCHEMATIC

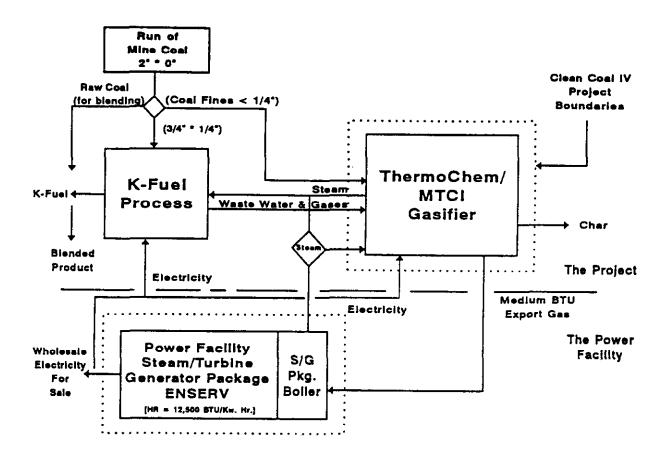


FIGURE 11: K-FUELS/THERMOCHEM COMMERCIAL PROJECT PROCESS STRUCTURE

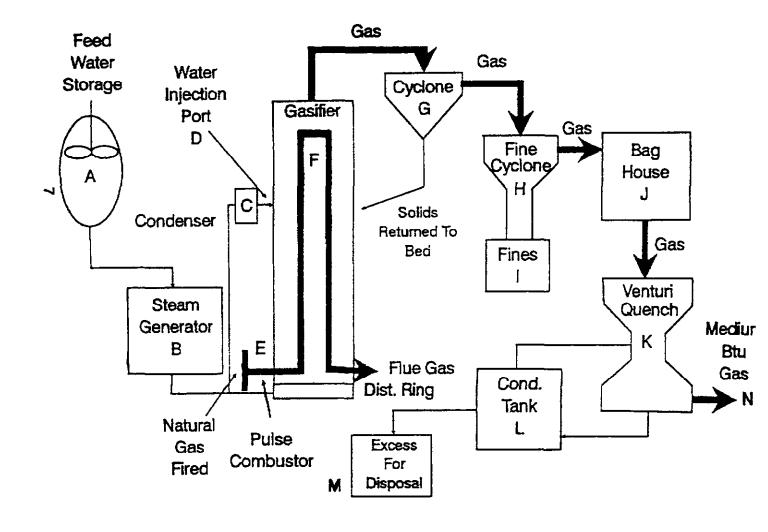


FIGURE 12: MTCI GASIFIER TESTS OF CABALLO ROJO COAL - DIAGRAM OF TEST FACILITIES

BLAST FURNACE GRANULAR COAL INJECTION

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ABSTRACT

A blast furnace coal injection system will be constructed and tested on large high productivity blast furnaces at the Burns Harbor plant of the Bethlehem Steel Corporation. This project will demonstrate injection facilities on two blast furnaces and will permit a comparison of operation with both granular (coarse) and pulverized (fine) coal injection. Injection rates up to 400 lbs/ton hot metal will be demonstrated with a variety of domestic coal types. With the completion of the National Environmental Policy Act (NEPA) process and issuance of a construction/air permit from the State of Indiana, the project has moved into the detailed design and construction stage with commissioning scheduled for early 1995.

INTRODUCTION

BACKGROUND - COAL INJECTION FOR BLAST FURNACES

Blast furnaces produce hot metal, which is used in the basic oxygen furnaces for refinement into various grades of steel. Major ingredients in the production of hot metal are iron ore, coke and limestone. As shown on Figure 1, the ironmaking blast furnace is at the heart of the integrated steelmaking process. Fine iron ore is agglomerated by pelletizing or sintering. The prepared ferrous materials, along with coke, are charged alone or in combination with lump iron ore into the blast furnace. Preheated air is injected near the bottom of the furnace and ferrous materials are reduced and melted by hot combustion products from the burning coke to produce molten iron. The molten iron is combined with scrap and flux and is refined in the steelmaking process. The basic oxygen furnace is the predominant method used in integrated steelmaking.

Figure 2 provides more detail on the blast furnace operation. As shown, the raw materials (ore, coke and limestone) are conveyed to the top of the furnace either on a conveyor belt or in a "skip" car. All or part of the limestone (and dolomite), which is used as flux to remove contaminants in the coke and ore, can be charged directly or combined in the ferrous sinter and pellet feed during their production.

The raw materials are charged to the top of the furnace through a lock hopper arrangement to prevent the escape of pressurized hot reducing gases. Air needed for the combustion of coke to generate the heat and reducing gases for the process is passed through stoves and heated to 1500-2300°F. The heated air (hot blast) is conveyed to a refractory-lined bustle pipe located around the perimeter of the furnace. The hot blast then enters the furnace through a series of ports (tuyeres) around and near the base of the furnace. The molten iron and slag are discharged through openings (tapholes) located below the tuyeres. Resultant molten iron flows to refractory-lined ladles for transport to the steelmaking shop.

A schematic showing the various zones inside the blast furnace is given on Figure 3. As can be seen, the raw materials, which are charged to the furnace in batches, create discrete layers of ore and coke. As the hot blast reacts with and consumes coke at the tuyere zone, the burden descends in the furnace resulting in a molten pool of iron flowing around unburned coke at the furnace bottom (bosh area). Reduction of the descending ore occurs by reaction with the rising hot reducing gas that is formed when coke is burned at the tuyeres.

The cohesive zone directly above the tuyeres is so called because it is in this area that the ore, which has been reduced is being melted and passes through layers of unburned coke. The coke layers provide the permeability needed for the hot gases to pass through this zone to the upper portion of the furnace. Unlike coal, coke has the qualities needed to retain its integrity in this region and is the reason that blast furnaces cannot be operated without coke in the burden.

The hot gas leaving the top of the furnace is cooled and cleaned. Since it has a significant heating value (80-100 BTU/scf), it is used to fire the hot blast stoves. The excess is used to generate steam and power and for other uses within the plant.

Over the years many injectants (natural gas, tar, oils, etc.) have been used in blast furnaces to reduce the amount of coke used. Their use is a matter of economics with each location making choices considering the site specific relative costs of coke and injectants available. Natural gas has been a common injectant used in this country. Recent technological developments in Europe and Asia, where coal has been widely used as an injectant, have established that the highest levels of injection and subsequent displacement of coke can be obtained by using coal.

A major consideration in evaluating coal injection in the United States is the aging capacity of existing cokemaking facilities and the high capital cost to rebuild these facilities to meet emission guidelines under the Clean Air Act Amendments. The increasingly stringent environmental regulations and the continuing decline in domestic cokemaking capability will cause significant reductions in the availability of commercial coke over the coming years. Due to this decline in availability and increase in operating and maintenance costs for domestic cokemaking facilities, commercial coke prices are projected to increase by more than general inflation. Higher levels of injectants, such as coal, enable domestic integrated steel producers to minimize their dependence on coke.

COAL PREPARATION AND INJECTION AT BURNS HARBOR

Natural gas is the injectant currently being used in the production of iron in the Burns Harbor blast furnaces of Bethlehem Steel Corporation. Even with maximum use of natural gas, the plant lacks sufficient cokemaking capability to support its ironmaking capability. That situation led Bethlehem to the decision to submit a proposal to the DOE to conduct a comprehensive assessment of coal injection in the Burns Harbor blast furnaces. The program is designed to provide the industry with comparative data on a variety of U.S. coal types, grind sizes, etc. Following an extensive review by the DOE, Bethlehem's Blast Furnace Granular Coal Injection System Demonstration Project was one of thirteen demonstration projects selected to enter into contract negotiations. During negotiations, the scope of the project was expanded to include improvements to the blast furnaces to enhance the potential for a successful demonstration.

The DOE financial assistance will enable Bethlehem to demonstrate and compare granular (coarse) coal injection with pulverized (fine) coal injection using a technology successfully employed by British Steel plc. Under the terms of the DOE financial assistance, Bethlehem will demonstrate both granular and pulverized coal injection at rates of up to 400 pounds per net ton of hot metal for a number of domestic coals.

PROJECT GOALS

As shown on Figure 4, this project will obtain comparative data for a variety of coal types, grinds and injection level. The primary thrust of the work is to demonstrate (a) conversion for, (b) optimization of and (c) commercial performance characteristics of granular coal as a supplemental fuel for steel industry blast furnaces. The technology will be demonstrated on large, hard-driven blast furnaces using a wide range of coal types available in the U.S. The planned tests will assess the impact of coal particle size distribution as well as chemistry on the amount of coal that can be injected effectively. Upon successful completion of the work, the results will provide to others the information and confidence needed to assess the technical and economic advantages of applying the technology to their own facilities.

TECHNOLOGY DESCRIPTION

Bethlehem's decision to utilize the Simon Macawber Blast Furnace Granular Coal Injection (BFGCI) System which can produce both granular and pulverized coal rather than a system which produces only pulverized coal (as has been more widely employed), is due to a variety of technical and economic advantages which made this system potentially very attractive for application in the U.S. basic steel industry. A schematic showing the application of the technology to the blast furnace is given on Figure 5. Following are some of the technical advantages associated with utilization of this system:

- 1. The injection system has been proven with granular coal as well as with pulverized coal. No other system has been utilized over this range of coal sizes.
- 2. The potential costs for granular coal systems are less than for pulverized.
- 3. Granular coal is easier to handle in pneumatic conveying systems. Granular coals are not as likely to stick to conveying pipes if moisture control is not adequately maintained.
- 4. Research tests conducted by British Steel indicate that granular coal is more easily maintained in the blast furnace raceway (combustion zone) and is less likely to pass through the coke bed. Coke replacement ratios obtained by British Steel have not been bettered in any worldwide installation.
- 5. Granular coal's coarseness delays gas evolution and temperature rise associated with coal combustion in the raceway. Consequently, it is less likely to generate high temperatures and gas flows at the furnace walls which result in high heat losses, more rapid refractory wear and poorer utilization of reducing gases.
- 6. System availability has exceeded 99 percent during several years of operation at British Steel.
- 7. High injection levels require accurate variable control of injection rates, both for individual tuyeres and the complete system. The unique variable speed, positive displacement Simon-Macawber injectors provide superior flow control and measurement over other coal injection systems.

HISTORY OF THE TECHNOLOGY

Coal injection into blast furnaces dates back more than 100 years; it was the first fuel known to have been injected. In the United States, pulverized coal has been injected into blast furnaces at the Ashland Kentucky Plant of Armco Steel since the mid-1960's. However, different economic situations at other facilities in the United States precluded wide application of coal injection technology. That situation has changed and a number of steel companies in the U.S. have installed or are planning to install coal injection facilities.

As with other companies, Bethlehem Steel has monitored the progress of blast furnace coal injection developments worldwide for a number of years. The development and application of a process that permits the use of granular (as well as pulverized) coal caught our interest. The equipment provides the capability of using either grind size, with the option of long-term use of the less expensive granular type.

The joint development between British Steel and Simon-Macawber for the injection of granular coal into blast furnaces began in 1982 on the Queen Mary Blast Furnace at the Scunthorpe Works. (1,2) The objective of the development work was to inject granular coal into the furnace and test the performance of the Simon-Macawber equipment with a wide range of coal sizes and specifications. Based on Queen Mary's performance, coal injection systems were installed on Scunthorpe's Queen Victoria, Queen Anne and Queen Bess (operational standby) blast furnaces and on Blast Furnaces 1 and 2 of the Ravenscraig Works. Queen Victoria's system was brought on line in November, 1984 and Queen Anne's in January, 1985. The Ravenscraig systems were started up in 1988. The success of the GCI systems at Scunthorpe and Ravenscraig, although demonstrated on smaller blast furnaces, led Bethlehem to conclude that the system could be applied successfully to large blast furnaces.

INSTALLATION DESCRIPTION

The coal preparation/injection facility will be retrofitted to blast furnaces, Units "C" and "D", at our Burns Harbor plant located in Porter County, Indiana, on the southeast shore of Lake Michigan. Highlights of the blast furnace and coal injection facilities are given on Figure 6. As noted on this Figure, Burns Harbor has experience with the injection of tar and oil as well as natural gas. This experience will be an asset when the coal injection trials begin.

A simplified flow diagram for the process is shown on Figure 7. The Raw Coal Handling Equipment and the Coal Preparation Facility includes the facilities and equipment utilized for the transportation and preparation of the coal from an existing railroad car dumper until it is prepared and stored prior to passage into the Coal Injection Facility; the Coal Injection Facility accepts the prepared coal and conveys it to the blast furnace tuyeres.

SITE LOCATION

The Coal Preparation Facility, the Coal Injection Facility and a utilities and control center for the facilities will be located within one building consisting of three attached structures. The building will be located between the two blast furnaces on a site currently occupied by a blast furnace warehouse and maintenance building which will be relocated. This location was chosen because it is the closest equidistant site to the two blast furnaces. Such location will minimize pressure drop and power requirements for transporting the coal to the blast furnaces.

RAW COAL HANDLING EQUIPMENT

Raw Coal Handling. Coal for this project will be transported by rail from coal mines to Burns Harbor similar to the way in which the plant now receives coal shipments for the coke ovens. The coal will be unloaded using an existing railroad car dumper, which is currently part of the blast furnace material handling system. A modification to the current conveyor will be made to enable the coal to reach either the coke ovens or the coal pile for use at the Coal Preparation Facility.

This modification will require a new 60-inch wide transfer conveyor to be installed from the existing conveyor and run east about 162 feet (40 feet above the ground) to a junction house. There the coal will be transferred to a new 60-inch wide stockpile conveyor which will run 760 feet to the north and end at the space for the new raw coal storage pile. The coal pile will be formed using a 200-ft. long radial stacker capable of building a 10-day storage pile (approximately 28,000 tons). The new material handling system from the car dumper to the coal storage pile will be sized at 2,300 tons per hour to match the output of the car dumper.

Raw Coal Reclaim. The raw coal reclaim tunnel will be installed underground beneath the coal storage pile. The concrete tunnel will be about 12 feet wide and 16 feet high and will contain three reclaim hoppers in the top of the tunnel. The reclaim hoppers, which are directly beneath the coal pile, will feed a 36-inch wide conveyor in the tunnel. The 500-ft. long reclaim conveyor will transport the coal at a rate of 400 tons per hour above ground to the south of the storage pile. A magnetic separator will be located at the tail end of the conveyor to remove tramp ferrous metals. The conveyor will discharge the coal onto a vibrating screen which will separate coal over 2 inches in size from the main stream of minus 2-inch coal. The oversized coal will vary depending on the weather (more during the winter when frozen lumps are expected) and will pass through a precrusher which will discharge minus 2-inch coal. The coal from the precrusher will join the coal that passed through the screen and will be conveyed from ground level by a 36-inch wide plant feed conveyor to the top of the building that houses the Coal Preparation Facility.

The reclaiming of coal from the pile will be done by gravity as long as there is coal above each of the reclaim hoppers. It will be necessary to have a bulldozer on the pile to push coal from the "dead" storage areas to the "live" storage areas above each of the reclaim hoppers.

COAL PREPARATION FACILITY

The plant feed conveyor will terminate about 95 feet high at the top of the building that houses the Coal Preparation Facility. Coal will be transferred to a distribution conveyor, which will enable the coal to be discharged into either of two steel raw coal storage silos. The raw coal silos will be cylindrical in shape with conical-shaped bottoms. They will be completely enclosed with a vent filter on top. Each silo will hold 250 tons of coal, which is a four-hour capacity at maximum injection levels. Air cannons will be located in the conical section to loosen the coal to assure that mass flow is attained through the silo.

Coal from each raw coal silo will flow into a feeder which controls the flow of coal to the coal preparation mill. In the preparation mill the coal will be ground to the desired particle size. Products of combustion from a natural gas fired burner will be mixed with recycled air from the downstream side of the process and will be swept through the mill grinding chamber. The air will lift the ground coal from the mill vertically through a classifier where oversized particles will be circulated back to the mill for further grinding. The proper sized particles will be carried away from the mill in a 52-inch pipe. During this transport phase, the coal will be dried to 1-1.5% moisture. The drying gas will be controlled to maintain oxygen levels below combustible levels.

The product coal will then be screened. Two full capacity parallel screens will be provided so that a screen can be changed without shutting down the coal preparation plant. The dried ground coal will be transported into one of four 180-ton product storage silos and will then be fed into a weigh hopper in one-ton batches. The one ton batches will be dumped from the weigh hopper into the distribution bins which are part of the Coal Injection Facility. There will be two grinding mill systems. Each system will produce 30 tons per hour of pulverized coal or 60 tons per hour of granular coal.

COAL INJECTION FACILITY

The Coal Injection Facility will include four distribution bins located under the weigh hoppers described above. Each distribution bin contains 14 conical-shaped pant legs. Each pant leg will feed an injector which allows small amounts of coal to pass continually to an injection line. Inside the injection line, the coal will be mixed with high-pressure air and will be carried through approximately 600 feet of 1-1/2-inch pipe to an injection lance mounted on one of the 28 tuyere blowpipes at each furnace. At the injection lance tip, the coal will be mixed with the hot blast and will be carried into the furnace raceway. The fourteen injectors at the bottom of the distribution bin will feed alternate furnace tuyeres.

Each furnace requires two parallel series of equipment, each containing one product coal silo, one weigh hopper, one distribution bin, 14 injectors, 14 injection lines and 14 injection lances.

TEST PLAN

The project will address a broad range of technical/economic issues as shown on Figure 8.

COAL GRIND SIZE

The project will evaluate coal injection over a broader range of coal particle sizes than has ever been conducted at any plant in the U.S. Only pulverized coal, defined as 70-80% minus 200 mesh (74 microns), has been injected commercially in the U.S. The primary focus of this project will be on granular coal, defined as 100% minus 4 mesh (5 mm), 98% minus 7 mesh (3 mm) and less than 30% minus 200 mesh (74 microns). The work will demonstrate on a commercial scale in the U.S. the coal preparation/injection system that can produce granular as well as pulverized coal. More important, it will show the effects of injected coal particle size on blast furnace performance. If the successful experiences of European operations with granular coal can be repeated or improved upon in the CCT III Project, then the advantages of granular coal over pulverized coal injection systems for commercial applications in the U.S. will have been demonstrated. These potential advantages include reduced capital cost for the grinding facilities and reduced consumption of electric energy (and other operating cost factors) for grinding the coal. The data to be generated on both fine and coarse injected coal will be of value in the planning of future U.S. commercial installations.

COAL INJECTION RATE

The plan for this project includes evaluating operations over a range of coal injection rates. We intend to push the upper boundaries of coal injection to 400 lbs of coal/NTHM. By operating and evaluating at coal injection rates ranging up to 400 lbs/NTHM, we will determine the technical limit for the coal injection system, establish the relationship between coal injection rate, furnace wall heat load, and any excessive wear of refractory lining to blast furnaces such as those at Burns Harbor; and confirm the operating costs and economic advantages that have been projected for coal injection.

COAL SOURCE

Our project will generate comparative data on coals with distinctly different chemical and physical characteristics. The plan is to use an Eastern bituminous coal with low ash and sulfur content; an Eastern bituminous coal with moderate ash and higher sulfur content; a Midwestern bituminous coal with higher inherent moisture but with low ash and moderate-to-high sulfur content; and a Western sub-bituminous coal with high inherent moisture but with low ash and sulfur content.

Each coal will be utilized for a sufficiently long period of time (about two months) to assess how it performs as a blast furnace injectant. Coal handling (i.e., grinding rates, injection system performance) and blast furnace parameters such as production, coke replacement, hot metal chemistry and slag volume are anticipated to be affected by the physical and chemical properties of the coal used for blast furnace coal injection. Data derived from this evaluation will make it possible for blast furnace operators to determine for themselves which coal would be most attractive for injection in their specific cases, including raw coal costs, transportation costs, coal grinding and injection costs, and the effects on blast furnace operations.

BLAST FURNACE CONVERSION METHOD

Neither of the two blast furnaces at Burns Harbor is equipped with coal injection facilities. In this project, we propose to convert both blast furnaces for coal injection during 1994. "C" Furnace is scheduled to be out of service for an extended reline in mid-late 1994. It is during this period that "C" Furnace will be fitted for coal injection. We propose to make the coal injection changes for "D" Furnace "on-the-fly", during very brief, perhaps eight hour outages. Thus, we will demonstrate the successful implementation of the modifications for blast furnace coal injection during both out-of-service and in-service modes. These will include planning and facilities for coal storage and handling, grinding, injection and alterations in the vicinity of the blast furnace itself (including work at the tuyeres).

Many of the physical components utilized in the coal injection system are also utilized in other commercial systems. The major portion of the technology envelope for this system is the integration of this equipment into a system that prepares coal as required for injection, allows flow to be controlled individually for each injection point into the blast furnace or allows all to be varied simultaneously, monitors the total amount injected and the flow to each tuyere, and includes the necessary know-how for injecting solid, granular fuel into a blast furnace. Key elements in this technology package are the weigh system, the variable flow injectors, lance sizing and positioning, and knowledge of how the factors of coal size, coal source and coal injection rate interact. Key elements of the portion of the project that pertain to blast furnace conversion methods involve the integration and coordination of engineering, construction and operations functions.

PROJECT SCOPE

To achieve these objectives, the demonstration project is divided into the three Phases (Figure 9).

Phase I - Design
Phase II - Construction
Phase III - Operation

At the present time, a turnkey contract has been placed with Fluor Daniel for the facility. Design Engineering is nearing completion. Equipment purchase orders have been placed with ATSI/Simon Macawber for the injection systems and site preparation is in progress. Regarding blast furnace improvements, those upgrades scheduled for the D furnace were completed during the last reline in late 1991. Planned major improvements to the C furnace will be completed during the reline of that furnace in the summer/fall 1994. The coal injection system is scheduled to be completed early in 1995 with testing to begin shortly thereafter.

REFERENCES

- D. S. Gathergood, "Coal Injection Into the Blast Furnace", International Iron & Steel Institute Committee on Technology, April 26, 1988.
- 2. D. S. Gathergood and G. Cooper, "Blast Furnace Injection Why Granular Coal"? Steel Technology International, 1988.

ELECTRICARC FURNACE OPEN HEAHTH FURNACE BASIC OXYGEN FURNACE DIRECT REDUCTION COKE OVENS PELLETIZEA C U

FIGURE 1
THE STEELMAKING PROCESS

CHECKERWORK BLAST COLD BRICK STACK STOVES COMBUSTION GAS CLEAN GAS GAS CLEANING, THE BLAST FURNACE COMPLEX HOT FURNACE GAS BLAST MOLTEN IRON FIGURE 2 BLAST FURNACE SLAG Ladle ČAST RAW MATERIAL CHARGING HOT CAR STOCK

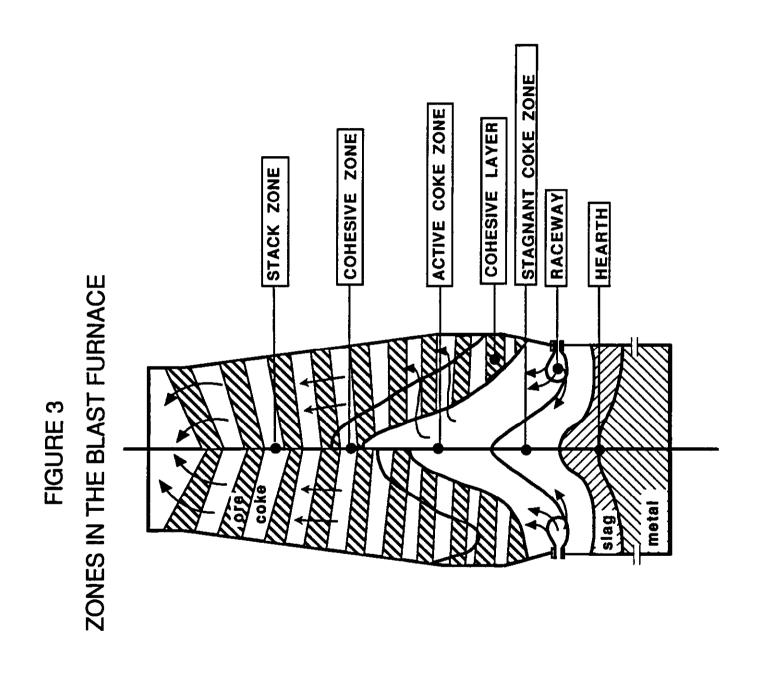


FIGURE 4

Coal Injection Test Program Parameters

Granular (100% -4 mesh) to Pulverized (75% -200 mesh)	Up to 400 lbs per NTHM	East, Midwest and West (Differing Chemical and Physical Characteristics)
•	•	ı
Coal Grind Size	Injection Level	Coal Types

Bethlehem Steel Corp.

System Installation

Reduced Coke

Requirement

During Furnace Reline and "On-the-Fly"

Less Reliance on Foreign Coke and/or Environmental Problems Associated

with Domestic Coke Production

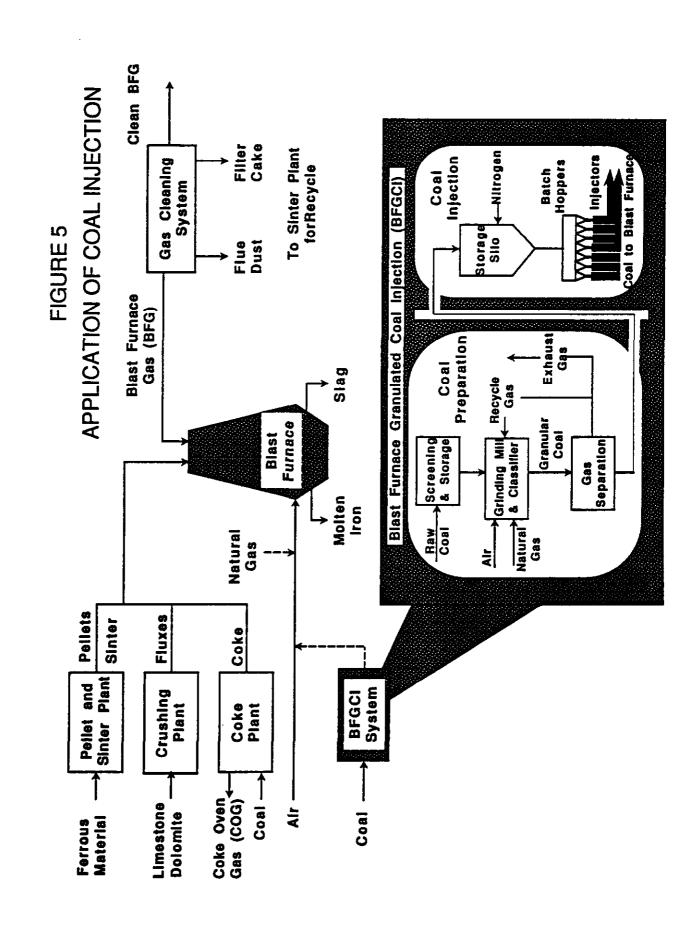


FIGURE 6

Coal Injection Test Site/Facilities

Bethlehem Steel Burns Harbor Plant, Porter County, Northern Indiana • Location

Blast Furnaces

Number

35 & 38 ft. Hearth Diameter

Approximately 7,000 tons/Day Pig Iron/Furnace (8 TPD per 100 cu.ft. Working Volume)

Natural Gas, Oil, Tar

Fuel Injection

Simon-Macawber

Coal Injection Facilities

Bethlehem Steel Corp.

Size

Production Rate

FIGURE 7 COAL INJECTION - BURNS HARBOR PLANT

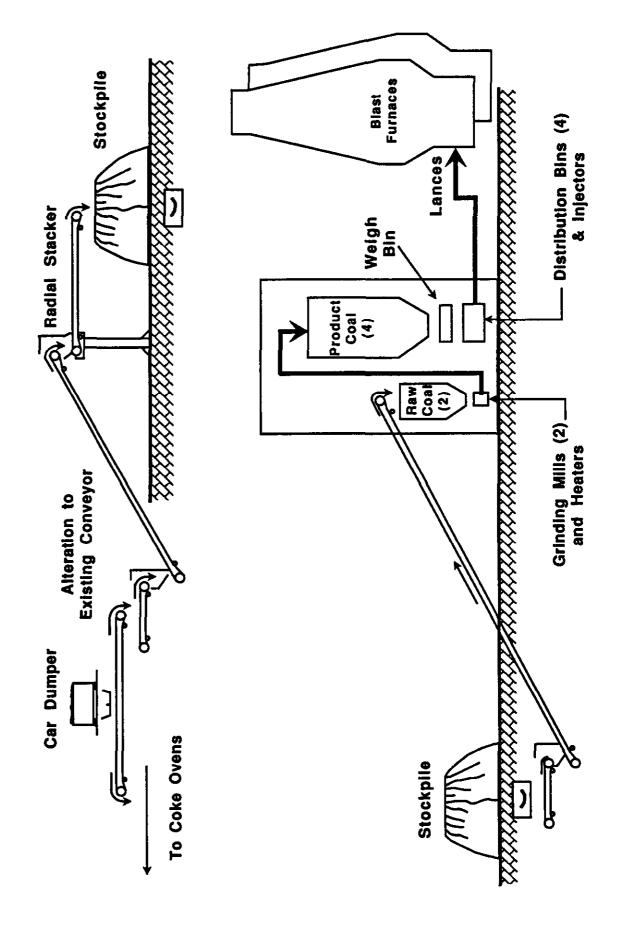


FIGURE 8

Coal Injection Test Program Goals

Granular (100% -4 mesh) to Pulverized (75% -200 mesh) Coal Grind Size

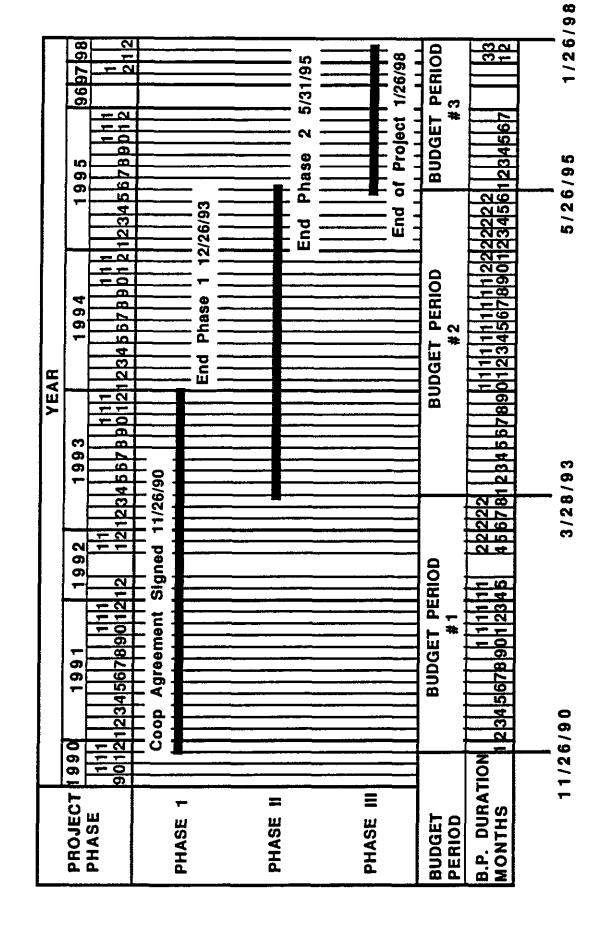
Up to 400 lbs per NTHM Injection Level East, Midwest and West (Differing Chemical and Physical Characteristics)

During Furnace Reline and "On-the-Fly" ı System Installation

Bethlehem Steel Corp.

Coal Types

FIGURE 9
PROJECT SCHEDULE



STATUS OF COAL TECH'S AIR COOLED SLAGGING COMBUSTOR-

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ABSTRACT

This paper summarizes the status of a six year development effort on a 20 MMBtu/hr slagging, coal combustor that was retrofitted to an oil designed package boiler. In addition to the efficiency benefits of regenerative air cooling, the combustor internally controls SO₂ and NO_x emissions. The combustor also substantially reduces dioxin emissions from coal and from coal cofired with refuse derived fuel. It has vitrified fly ash containing a wide range of unburned carbon. To date, the combustor has operated for about 1600 hours, with about one-half of this time on coal, and the balance on oil and gas. Current test efforts are focused on automatic computer control of the combustor in order to demonstrate its durability in continuous coal fired operation. In addition, systems and cost analyses have been performed on applications of the combustor to retrofit and repower industrial boilers and combined gas turbine-steam turbine power plants. Installed retrofit costs for the combustor are estimated at under \$10/lb of steam for industrial boilers, and from \$86/kW for small power plants to \$172/kW for a 250 MW power plant. The estimated cost of a 20 MW greenfield combined cycle plant system is in the \$1200 to \$1400/kW range.

INTRODUCTION

This paper summarizes the status of Coal Tech's commercial scale demonstration of a patented air cooled, slagging coal combustor. Air cooling recycles the combustor wall heat transfer loss to the combustion air, which makes it available to the peak of the thermodynamic cycle. On the other hand, water cooling of the combustor yields low temperature heat which is difficult to utilize in a thermodynamic cycle. Typically, the heat losses to the water cooled sections of this combustor are between 2 and 3%, compared to 8 to 10% if the entire combustor were water cooled. By proper combustor design, the energy needed to drive the cooling air can be as little as 1 to 2% of the total heat input. Therefore, the overall efficiency loss in an air cooled combustor can be as little as one-half that of the water cooled combustor. A portion of the SO₂ and NO_x emissions are controlled inside the combustor. The combustor is designed for new and retrofit boiler applications. The air cooled combustor development began in the late 1970's using a 1 MMBtu/hr air cooled cyclone combustor [1]. Development continued in the mid 1980's with SO₂ and NO_x control tests in a 7 MMBtu/hr water cooled cyclone combustor [2]. This work was followed by the design, construction, and installation of the present 20 MMBtu/hr, air cooled, combustor between 1984 and 1987 [3]. The combustor was first tested in 1987 with coal water slurry fuels, and then converted to pulverized coal operation.

The first three years of the demonstration effort were conducted under DOE Clean Coal Program sponsorship. During the Clean Coal project, which began in 1987, many of the operational issues involved in using an air cooled combustor were resolved during nearly 800 hours of combustor operation. About 1/3 of the test hours were on coal.

Since the completion of the Clean Coal tests, the combustor has been used on other test projects. Tests were conducted on ash vitrification [10] and refuse derived fuel combustion [15]. During these tests, the data base developed during the manually controlled Clean Coal combustor tests was used to automate the combustor's operation. For this purpose, a process control software was specialized for the combustor's operation and installed on a micro-computer. In addition, major progress was made on improving the combustion efficiency, SO₂ reductions, reliability, and durability.

Current DOE sponsored tests focus on round-the-clock, coal fired operation under automatic computer control. The objectives are to acquire a data base on durability of combustor components, durability of the auxiliary components needed to operate the combustor, and on the impact of the combustor on the boiler efficiency, fouling and corrosion. Another key objective is

to remove essentially all of the coal sulfur in the combustor with sorbent injection. Finally, the application of the combustor to a wide range of end uses, such as the retrofit and repowering of industrial boilers and power plants, combined cycle industrial power plants, cofiring of coal and waste fuels, firing low grade high ash coals, and vitrifying high carbon content fly ash, is being investigated.

Progress reports on the air cooled combustor tests were presented at the 5th Annual Pittsburgh Coal Conference [4] in September 1988, the 82nd Air Pollution Conference [5] in June 1989, and the 7th Annual Pittsburgh Coal Conference in September 1990 [6]. The economics of emission control in utility boilers with this combustor were first presented in March 1990 [7]. A detailed report on the Clean Coal Project was published in August 1991 [8]. More detailed descriptions of the work described in this paper were recently reported elsewhere [12,17,18,19,24,27]. Due to recent progress in the development effort, there have been significant improvements in the combustor performance and in the design of the combustor-boiler system. These design changes have substantially lowered the projected installed combustor cost from previously reported levels. Designs have been developed for combustors rated up to 150 MMBtu/hr for application to boiler retrofit and to new boilers whose design is integrated with the combustor.

Coal Tech's Advanced Air Cooled, Cyclone Coal Combustor

The cyclone combustor is a high temperature (> 3000°F) device in which a high velocity swirling gas is used to burn crushed or pulverized coal. The ash is separated from the coal in liquid form on the cyclone combustor walls, from which it flows by gravity toward a port located at the downstream end of the device. A brief description of the operation of Coal Tech's patented, air cooled combustor is as follows (see Figure 1): A gas and oil burner, located at the center of the closed end of the unit, is used as a pilot to pre-heat the combustor and boiler during startup. Dry pulverized coal and sorbent powder for SO₂ control are injected into the combustor in an annular region enclosing the gas/oil burners. Air cooling is accomplished by using a ceramic liner, which is cooled by the swirling secondary air. The liner is maintained at a temperature high enough to keep the slag in a liquid, free flowing state. The liquid slag is drained through a tap located at the downstream end of the combustor.

Nitrogen oxide emissions are reduced by operating the combustor fuel rich. Between 67% and 80% NO_X reductions were measured in pilot combustors rated at 1 MMBtu/hr [9] and at 7 MMBtu/hr [10]. In the 20 MMBtu/hr combustor, about two-thirds stack NO_X reductions to less than 200 ppm (normalized to 3 % O₂) have been measured under staged operation with

combustion efficiencies of 95% to 99%. Efficient combustion under fuel rich conditions requires either uniform solids feed or combustion gas temperatures in the 3400°F range. With feed non-uniformities and gas temperatures in the 3000 to 3200°F range, the measured combustion efficiencies in the 20 MMBtu/hr air cooled combustor averaged around 85% at a 0.7 stoichiometric ratio. At this condition, NO_X emissions are reduced to only 350 ppm (at 3% O₂), or about 33% below excess air levels.

A major focus in the air cooled combustor's development was the control of sulfur emissions by means of Coal Tech's patented, sorbent injection process into the combustor. The process is based on non-equilibrium chemical capture of the sulfur by the sorbent particles during the 0.1 second gas transit time in the combustor. The sulfur bearing sorbent particles can exit the combustor with the combustion gas into the lower (<2000°F) temperature zone in the boiler before the reaction reverses itself. Alternatively, the sulfur bearing sorbent particle can impact and dissolve in the slag and exit from the combustor before the reaction reverses itself. To retain the sulfur in the slag, the liquid slag transit time in the combustor must be less than several minutes. This is difficult to achieve, and to date, the highest sulfur concentration measured in the 20 MMBtu/hr combustor has been 20% of the coal sulfur. On the other hand previous results obtained in the 7 MMBtu/hr combustor tests [10] yielded SO₂ reductions approaching 100% [measured at the stack exhaust] with limestone injection in the first stage. After extensive testing, during the past year, SO₂ reductions in the 85% to 90% range were measured at the stack using calcium hydrate injected into the 20 MMBtu/hr combustor at a Ca/S mol ratio of 3 to 4. Testing is in progress to determine the relative magnitude of sulfur capture in the combustor and boiler due to sorbent injection in the combustor. Recent sulfur capture results will be summarized in this paper.

Description of the 20 MMBtu/hr Combustor-Boiler Test Facility

The design of the 20 MMBtu/hr Coal Tech combustor is based on the detailed design of an air cooled combustor at thermal input ratings of 100 MMBtu/hr [11]. The latter size was initially selected because it was the most probable market size for this combustor. The 20 MMBtu/hr combustor was initially selected for application with coal water slurry fuels, and subsequently for commercial applications to small industrial boilers. The 20 MMBtu/hr combustor was installed on a 17,500 lb/hr steam boiler in an industrial plant in Williamsport, PA in early 1987. Figure 2 shows a side view drawing of the combustor attached to the boiler. The coal is pulverized offsite, and stored in a 4 ton capacity coal storage bin next to the boiler house. The coal is metered and fed into a pneumatic line to the combustor. The bin is refilled from a 24 ton trailer parked

outside the boilerhouse without combustor shutdown. Since the combustor's best slag retention is in the 70% to 80% range, it does not meet local particulate emission standards of 0.4 lb/MMBtu. Therefore, a wet particulate scrubber is used for this purpose. Slag drains from the combustor through an opening at the downstream end of the combustor (See figures 1 and 2) into a water filled tank. The slag is removed from the tank by means of a mechanical conveyor and deposited in a drum. The fuel and air streams to the combustor are computer controlled using the combustor's thermal performance as input variables. Diagnostics consist of measurement of fuel, air and cooling water flows, combustor wall temperatures, and stack gas measurements, including O2, CO2, CO, SO2, NOx, and HC. Gas samples are taken in the stack above the boiler and in the exhaust from the wet scrubber. Gas samples are also taken at the exhaust from the combustor into the boiler with a water cooled probe that is inserted through the rear boiler wall.

TEST RESULTS

Test Activities Dealing with the Combustor's Operation

A systems approach has been taken to the development of the combustor because auxiliary subsystems, such as coal feed, sorbent feed, combustion air supply, slag removal from the combustor, ash control in the boiler, and the combustor-boiler interface, directly impact the combustion efficiency, environmental control, and durability of the combustor. For example, high combustion efficiency and substantial SO₂ reductions were achieved only after a method for uniform coal and sorbent feed into the combustor was developed. Another area of extensive development was on the method to remove liquid slag from the combustor. A decrease of only several hundred degrees Fahrenheit in the slag temperature increases its viscosity to the point where slag flow ceases. Therefore, designs and procedures had to be developed which would maintain liquid slag flow in the combustor, and to clear the frozen slag that periodically accumulated in the slag tap. These consisted of adding local heaters to the slag tap section and adding an automated mechanical device that periodically breaks loose accumulated frozen slag from the slag tap.

In the first years of the present test effort, the combustor was operated under manual control. These tests showed that continuous real time control of the combustor's operation is very critical for durability, efficient combustion, and environmental performance. This control is critical with air cooled combustor walls because wall materials can rapidly degrade with wall temperature excursions. Therefore, beginning in 1990, a computer based control system was developed which allows completely automatic operation of the combustor. With computer control, it has been possible to replenish the ceramic walls of the combustor with frozen coal slag, essentially

eliminating the need for periodic patching of the ceramic wall material. For this procedure to function properly, it is essential to maintain the ceramic liner-combustion gas interface at a constant temperature of about 2000°F, within a variation of about 50°F. This degree of wall temperature control has been recently achieved in continuous combustor operation tests, each of which extended over 24 hours. No refurbishment of the refractory lined combustor wall was required between these tests. Tests of longer continuous operation are planned in the near future.

To date about 1600 hours of operation have been accumulated. In the course of testing, design improvements to the combustor and boiler system were installed and tested. For example, the 20 MMBtu/hr combustor was originally designed for cyclic operation with daytime coal firing and nighttime shutdown or pilot gas heat input operation. As a result, certain components, such as the combustor-boiler interface section, were not designed for round-the-clock coal fired operation at peak rated heat input. In the current test effort, these components were redesigned and tested for round the clock operation. Round the clock operation at steady heat inputs were recently implemented with scheduled 24 hour periods of continuous operation at 14 to 19 MMBtu/hr with 10 hours of coal firing, followed by 10 hours of No.2 oil, followed by 3 to 4 hours on coal. Post test evaluation of the combustor revealed no degradation of the combustor's internal wall. As a result, longer duration test will be implemented shortly.

An important element of the combustor test effort is analytical computer modeling to develop scaling relationships by comparing the modeling results with combustor test results. A two dimensional combustion code developed at Brigham Young University [16] is being used for this purpose. This code follows a set of coal particles that represent a typical coal size distribution from injection to final burnup or exit from the combustor. The modeling will be used to optimize the combustor's solids injection geometry and length to diameter ratio for a range of thermal inputs. Initial results are in the process of being analyzed...

Finally, the test effort yielded design improvements which simplify the combustor's fabrication and enhance its performance. As part of this performance enhancement, the air cooling and combustion air flow paths were redesigned to reduce the parasitic power that is required to drive the fans. These modifications have been recently incorporated in the design of a series of combustors ranging from 40 to 150 MMBtu/hr, whose installed cost is lower than the costs estimated from the current design. The costs given in the system section of this paper are based on these new designs.

Environmental Performance

(i) SO₂ Emissions

Sulfur capture by injected sorbents in the combustor is a non-equilibrium process. The gas residence time in the combustor is short, typically about 100 to 200 milliseconds [20,21]. A theory to fully explain all these effects has not yet been developed. The authors believe that the wide variability in SO₂ reduction data with combustor sorbent injection is due to variation in operating conditions.[22].

The following is a summary of the SO₂ reduction results in the 20 MMBtu/hr combustor: Initial results showed considerable variability due to non-uniform conditions. After major improvements in combustor performance were achieved in the past two years, especially in the area of feed uniformity, limestone injection yielded reductions of 56% at a Ca/S ratio of 2. Calcium hydrate injection in the combustor yielded SO₂ reductions in the range of 85% at Ca/S ratios somewhat greater than 3. All these measurements were obtained in the stack of the boiler, and as was recently verified some of this sulfur reduction took place inside the boiler. This will be discussed in the next paragraph. These reductions are based on the coal sulfur content. While the main controlling parameters have been identified, and SO₂ reductions as high as 90% have been measured in recent tests, past experience suggests that until this result is repeated numerous times under identical conditions, some uncertainty remains whether all the governing parameters have been identified.

During the past year, the emphasis on combustor tests has been on automatic operation and durability. SO₂ emissions have been measured in each test. Figure 3 shows a statistical average for all the tests of the past year of the SO₂ reduction measured at the boiler outlet as a function of the total Ca/S mol ratio. In the tests, calcium hydrate was injected for sulfur capture and an equal quantity of limestone was injected to improve slagging in the combustor. As noted above, limestone has been observed to be between 2 to 3 times less effective than calcium hydrate in capturing sulfur in the combustor. Therefore, the combined Ca/S mol ratio shown in figure 3 was about 1.3 times greater than would be required only with calcium hydrate. Despite non-optimized conditions, 70% reduction of SO₂ has been measured as a Ca/S of 4. This is equal to a Ca/S of 3 when the effectiveness of limestone is normalized to that of calcium hydrate.

To identify the relative degree of sulfur capture in the combustor and boiler with combustor injection of sorbent, gas samples were obtained inside the boiler by placing a probe within several

feet of the exhaust region from the combustor. Here the gas temperature is in the 2000°F to 3000°F range. For this one test, the SO₂ reduction due to sorbent injection in the combustor was 19% at the combustor exit and 48% at the boiler outlet, namely at the base of the stack. This was the first direct confirmation that the sorbent continues to react substantially in the furnace and convective sections of the boiler. As these measurements are repeated in future tests, parametric data on the relative effectiveness of sulfur capture in the combustor and boiler will be obtained. Finally, as noted in the Introduction, a maximum of 20% of captured sulfur was measured in the slag removed from the combustor. It is planned to focus the tests on optimization of sulfur capture with sorbent injection in the combustor after the automation and durability tests are complete.

(ii) Fly Ash Vitrification and Solid Waste Disposal

Beginning in 1988, several dozen combustor tests were performed on fly ash vitrification. Ash injection rates up to 55% of the combined ash-coal flow were achieved. Slag samples were unreactive as per the EPA Reactivity Tests for sulfides and cyanides. The trace metal leachate levels were within the EPA Drinking Water Standard. Slag chemical analysis and other properties indicate that the material is not classified as a hazardous waste. Detailed discussion of trace metal behavior in the combustor is given elsewhere [10]..

One important application of the combustor is for the conversion of high carbon content fly ash into vitrified slag. This type of ash has been found in the exhaust of pulverized coal fired boilers that have been converted to low NO_X coal burners. Recently, a test was performed with such a fly ash in which the carbon content was 30%. The ash was cofired with oil in order to obtain an accurate mass balance. In commercial use, coal would be used as the auxiliary fuel. The result showed that the slag produced in this test had no detectable carbon. From the carbon content of the fly ash that escaped the combustor and was captured in the stack particulate scrubber, it was determined that the carbon content of the original fly ash was reduced from 30% to 4.5%. An average of 85% of the carbon was found to be consumed in the combustor. The total quantity of injected fly ash was 200 pounds in a little over one hour. This was too small a quantity to perform a mass balance in order to determine the amount of slag conversion in the combustor.

Based on these results, it was determined that the cost of using the air cooled slagging combustor to vitrify a 30% carbon content fly ash from an 80 MW power plant could be recovered in about 1 year from the savings in eliminating fly ash disposal and lost heating value.

(iii) Air Toxics

The emissions of organic micropollutants from fossil fuel combustion sources is a matter of increasing importance. In 1990, a series of tests on refuse derived fuel (RDF) combustion were performed in the 20 MMBtu/hr combustor. As part of this test effort, the magnitude of organic micropollutants was measured in the stack. The RDF was cofired with coal, in various ratios up to 33% by weight of RDF. To provide a baseline for these tests, the stack micropollutants were also measured with only coal firing. Three classes of organics were measured: dioxin and furans, (PCDD, PCDF, {polychlorodibenzodioxins/polychlorodibenzofurans}) and PAH (polycyclic aromatic hydrocarbons). The dioxin compounds range from the tetra dioxins (TeCDD) to the octa-congeners (OCDD). The former are 1000 times more toxic than the latter. Measurements were taken inside the boiler and in the stack. Detailed results of the sample analyses are reported elsewhere [23,24].. The average level of PCDDs for coal only firing as measured at the stack was 22.5 ng/Nm³, and the PCDF levels at the stack were 7 ng/Nm³, both at 7%O2. For the cofired RDF-coal case, the corresponding levels were 1457 ng/Nm³ and 28 ng/Nm³. The first number is in the mid range of emissions from municipal incinerators [28]. However, the most toxic TeCDD's were only 10.3 ng/Nm³, or 0.7% of the total 1457 ng/Nm³ PCDD emissions in the coal-RDF case, and they were below the detection limit with coal only. Also, it is important to note that due to a temperature limitation problem with the probe used for this stack sampling, it was necessary to operate the combustor at high excess air conditions in the final burnup stage in the boiler. As a result, the CO level in the stack approached 1000 ppm, which was about 10 times greater than under normal coal firing. It is thus most probable that the level of the PCCD and PCDF emissions from RDF would be much reduced under optimum burnup conditions.

APPLICATIONS OF THE AIR COOLED SLAGGING COMBUSTOR

Use of the Combustor in a Combined Cycle Power Plant

The combustor can be used with a wide range of fuels, including pulverized coal, shredded refuse derived fuels, oil, sludge waste fuels, or natural gas. The use of air cooling makes the combustor attractive for integration into a combined gas-steam turbine power cycle. The exhaust of a natural gas or oil fired gas turbine contains sufficient oxygen and its temperature is in a suitable range for use as pre-heated combustion air in the combustor. The combustor is attached to a boiler which drives a steam turbine. Part of the steam is extracted from the turbine in order to augment the gas turbine power output with steam injection.

There are several cycle configurations that can be analyzed, depending on the ratio of gas turbine to steam turbine power output. To achieve maximum efficiency, this ratio should be greater than 50%, i.e. the gas turbine power being at the high temperature end of the cycle should be maximized. However, this cycle would require either high cost, natural gas for over 50% of its fuel input, or a high capital cost, coal gasifier for the gas turbine. The much lower cost slagging combustor cannot be used to fire the gas turbine.

For these reasons, a cycle was selected which maximizes the benefit of the combustor, although it yields a lower cycle efficiency. To quantify the thermodynamic and economic analysis, a nominal 20 MW combined cycle plant was selected in which the gas turbine produced about 25% of the power while the steam turbine produced the balance. Figure 4 shows a schematic of this combined power cycle. The base case consists of a commercial natural gas fired turbine operating at a nominal 1800°F turbine inlet temperature [29]. Its rated output is 5,940 kW with steam injection. The gas turbine exhaust steam provides the combustion air for the coal fired, air cooled combustor. In the 20 MW power plant, there are two combustors, each of which is attached to a separate factory assembled industrial boiler. Each of the two boilers produces 63,000 lb/hr superheated steam at 900°F, 950 psi. The steam drives a 13,200 kW turbine-generator. The steam turbine has two extraction points, one provides the steam for injection into the gas turbine, while the other (not shown in figure 4) is used for feedwater heating. The balance of the steam goes to the condenser. This arrangement yields about 25% of the power output from the gas turbine, with the balance provided by the steam turbine. The plant has a cycle efficiency of 32.48% with the commercial 1800°F gas turbine. With an advanced gas turbine having an inlet temperature of 2300°F, the cycle efficiency increases to 34.5%.

A plant layout and cost estimation analysis of the 20 MW power plant was performed. With the exception of the air cooled coal combustor, all other major components are commercially available. Budgetary vendor quotations for all major components and sub-systems were obtained. The total cost of this greenfield plant was \$24 million for about 19,000 kW, or \$1265/kW. This compares with a cost of \$1400-\$1750/kW for natural gas fired Cheng combined cycle [30] and a cost of \$2000-\$2300 for a fluid bed combustion, steam cycle [31].

Application of the Combustor to a 250 MW Power Plant.

The economics of retrofitting Coal Tech combustors to a 250 MW coal fired plant were analyzed using the procedures recommended by DOE for evaluating Clean Coal technologies [12]. This consists of applying a process contingency and a retrofit difficulty factor to the installed cost of

the new equipment added to an existing 250 MW coal fired plant. The added equipment consisted of a sorbent storage and feed system, sixteen Coal Tech air cooled coal combustors, and a slag removal system. Details of the procedures used for this analysis are given in reference 7. For the present paper, the economic analysis was updated by using the current combustor design for estimating the cost of each 150 MMBtu/hr combustors. The installed combustor cost was increased by a factor 1.94 for the contingency factors, and the cost of the other components, which are commercial, were increased by a factor of 1.1. Environmental performance data based on the best results achieved to date, namely, NO_x reductions of 80% and SO₂ reductions of 90%, with only combustor sorbent injection, were used in the analysis. The total capital cost for the retrofit was \$43 million, when the other cost factors listed in reference 12 are added to the process equipment capital cost. This cost equals \$172/kW.

Since the purpose of the retrofit is to reduce SO₂ and NO_x emissions, the conversion cost analysis was structured to allow a determination of the incremental cost of meeting these requirements. The analyses of the operating and maintenance items using the procedures and consumable costs of reference 12 showed that the variable operating costs were the largest contributor to the total operating costs. The sorbent, either limestone or calcium hydrate, each at a Ca/S mol ratio of 3, was the largest contributor to the variable operating costs. Parasitic power requirements to operate the combustors were a smaller, but still a substantial contributor. Using limestone, 15 year levelized operating costs were 7.36 mills/kW-hr and 8.01 mills/kW-hr for 2.5%S and 4.3%S coals, respectively. With calcium hydrate, the 15 year levelized cost increases to 9.23 mills/kW-hr for the 2.5% sulfur coal. This analysis assumed a 25%-75% equity-debt ratio with a 10% cost of funds and a 10% opportunity cost. These operating costs are about 30% less than the values quoted in the EPA/EPRI study¹⁴ for 10 different LIMB cases, and they are less than one-half of the equivalent wet flue gas scrubber costs. The economic assumptions used in reference 14 are not known to the authors. Based on the capital costs listed in reference 14, they could not have differed significantly for the present values.

With limestone, the 15 year levelized cost of retrofitting the 250 MW power plant with the combustor yields a cost of \$308/ton of SO_2 and NO_X with 2.5% sulfur coal. For 4.3% sulfur coal, the cost is \$197/ton. The unit cost decreases with increasing coal sulfur content because the capital costs are essentially independent of sulfur content.

Application of the Combustor to the Retrofit of a 120,000 lb/hr Coal Fired Boiler

A recent analysis was performed to convert a pair of 120,000 lb/hr industrial coal fired boilers with the air cooled combustor. The installed cost of the conversion was less than \$10/lb of steam, i.e. \$2.4 million. This cost was obtained from budgetary vendor quotations for the fabrication of the combustors, all the combustor auxiliary components, the combustor instrumentation and controls, and the installation of the combustors on the boilers. Since the use of this combustor allows selection of a lower grade high ash coal as a fuel, the potential fuel saving alone is sufficient to recover the conversion cost within two to three years. In addition, in the particular boiler under consideration, the present combustion efficiency was poor due to the design of the furnace section. Adding the fuel savings from the high combustion efficiency in the slagging combustor reduces the cost recovery to a one to two year period.

Application of the Combustor to Retrofit & Repowering of 20 MW Power Plants.

Using the same economics as in the previous sub-section, a cost of \$86/kW was obtained for the retrofit of a coal fired boiler with the air cooled combustor in a 20 MW power plant. In this case, the only new equipment consisted of the combustor, auxiliary combustor components such as a blower, pumps, valves, combustor controls and instrumentation, and combustor installation on an existing boiler.

Another site specific application that was investigated was the repowering of a 20 MW power plant with the air cooled combustor. In this case, the added equipment consisted of a coal pulverization and feed system, a limestone storage and feed system, an oil storage and feed system, a boiler, a slag removal system, a system for fly ash reinjection into the combustor from the baghouse, a baghouse, a stack, and a boilerhouse and associated structures. The existing turbine-generator, feedwater heating, and power transmission system would be refurbished. The estimated installed cost, using budgetary vendor quotations, was \$650/kW. A blended fuel would be used consisting of 75% (by weight) of a high ash coal waste, 20% bituminous coal, and 5% number two oil, with a combined cost of \$0.66/MMBtu. Income is derived from power sales to a regional electric utility for a 10 year period. The economic analysis used 20% equity, 80% debt financing at a 7.5% interest rate, seven year amortization, and a 40% tax rate. This yielded an attractive internal rate of return on equity of 28%. Other rate of returns can be derived by varying these economic assumptions.

Application to High Carbon Content Fly Ash Vitrification

This application was discussed in the "Test Results" section of this paper. The 30% carbon content of the fly ash tested is being produced in an 80 MW power plant at the rate of 6 tons/hour. A single slagging combustor can vitrify this ash and burn its carbon with the addition of coal and sorbent. The economics of the vitrification are very site specific. They depend on the carbon content of the ash, the ash disposal costs, the power production costs, and the market value of the slag. For the 80 MW plant studied, the increased combustion efficiency from carbon recovery in the fly ash and from elimination of fly ash disposal, allows recovery of the cost of the slagging combustor installation in less than 1 year.

CONCLUSIONS

The present six year test effort is the first commercial scale demonstration of this air cooled, slagging coal combustor. The initial three year test effort provided an operational data base for the combustor. These data have been subsequently incorporated in an automatic computer controlled combustor operating system which has substantially improved its performance, its environmental control, its reliability, and the durability of the refractory combustor wall. Wall durability requires maintaining the internal wall temperature in the 2000°F range to within 2% to 3%. This has been recently accomplished by using computer control for several continuous periods of 24 hour duration, without refurbishing the combustor wall between test periods.

Peak SO₂ reductions in the 85% to 90% range have been measured in the stack with calcium hydrate injection into the combustor. NO_x reductions in the 67% range have been measured in the stack with fuel rich combustor operation. The slag removed from the combustor is chemically inert. Cofiring of coal and refuse derived fuel in the combustor has yielded substantial reductions in the emissions of organic micropollutants.

The combustor was analyzed for various application, including a new 20 MW combined gassteam turbine power plant, retrofit to a 250 MW coal fired power plant, repowering of a 20 MW power plant, retrofit of industrial boilers, and fly ash vitrification. In all cases the combustor offers significant performance and cost advantages over competitive technologies.

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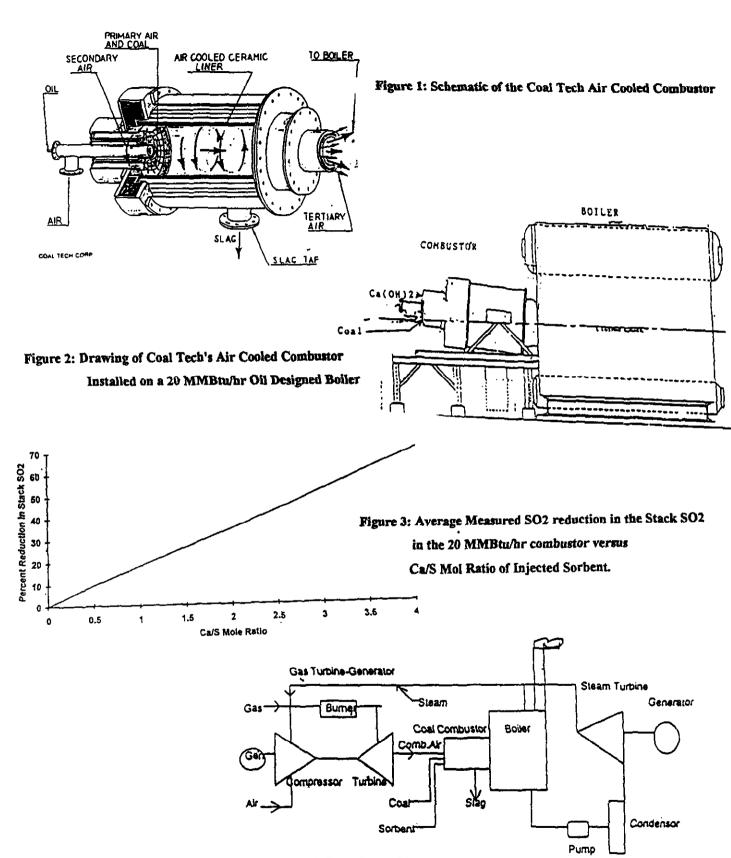


Figure 4. Schematic of An Industrial Combined Cycle Power Plant
Using the Coal Tech Combustor